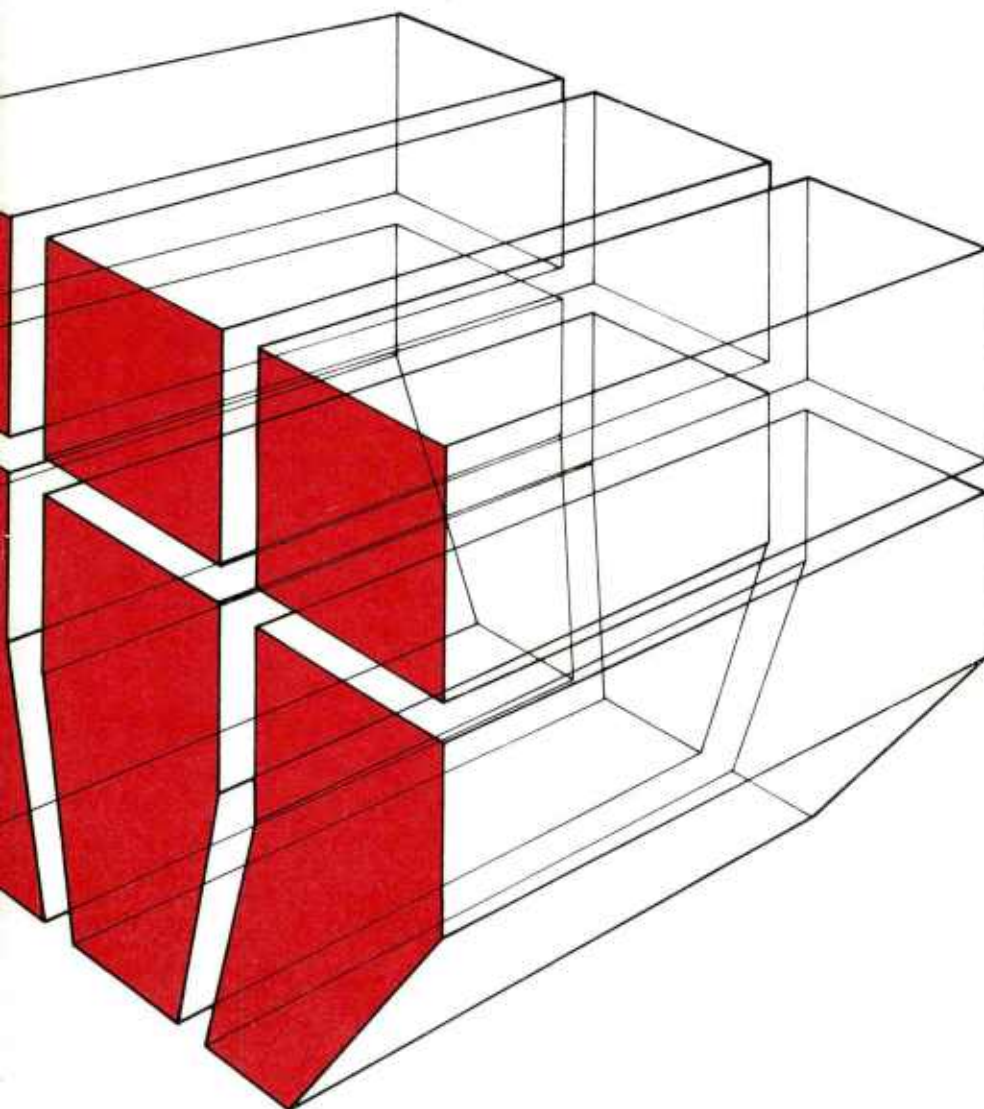**construction
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TECHNICAL REPORT N-93
September 1980

ANALYSIS OF FLUE GAS DESULFURIZATION (FGD)
PROCESSES FOR POTENTIAL USE ON
ARMY COAL-FIRED BOILERS

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by
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This report analyzes and evaluates existing flue gas desulfurization (FGD) techniques which could be used on Army boilers. Only those processes which are proven (in use for two or more years) or which have potential for use on Army-sized boilers in the near future are considered. In addition, this report presents guidelines to help Army installations select FGD processes appropriate to their needs.

Block 20 continued.

The report provides detailed economic, technical, and historical data on the single alkali, limestone, lime, double alkali, citrate, and magnesia FGD processes. Information on FGD capital costs and operation and maintenance cost is presented for boilers ranging from 25 to 250 x 10^6 Btu/hr (7.33 to 73.27 MW) heat input in size.

This report recommends three FGD systems for use in Army coal-burning boilers: single-alkali scrubbing, limestone scrubbing, and lime scrubbing.

SUMMARY

Three FGD systems are recommended for coal-burning boilers at Army installations. Many other systems are available, but single-alkali scrubbing, limestone scrubbing, and lime scrubbing are preferable because of their effectiveness, simplicity, reliability, operational history, minimal energy requirements, and flexibility.

Single-alkali flue gas scrubbers are gas-liquid contacting devices that use the chemical reactions between soluble alkali -- such as $NaOH$ or Na_2CO_3 -- and SO_2 to remove the oxides of sulfur from the combustion gases generated during the operation of fossil-fired furnaces.

Advantages of the single-alkali scrubbing process include its ease of operation, the highly reactive agent used, and its freedom from plugging and scaling. The process also does not produce any solid by-product and can tolerate some fly ash in the system. However, the process has some disadvantages -- especially waste disposal. Since sodium salts are soluble, the liquid waste must usually be treated before it can be discharged to the environment. The cost of reagents used, whether $NaOH$ or Na_2CO_3, is also frequently much higher than SO_2 scrubbing reagents such as lime and limestone. The use of alkali, especially caustic ($NaOH$), also requires increased concern with safety procedures for operating personnel.

Limestone flue gas scrubbers are gas-liquid contacting devices that use the chemical reactions between limestone (mostly $CaCO_3$) and SO_2 to remove the oxides of sulfur from the combustion gases generated during the operation of fossil-fired furnaces.

Wet limestone scrubbing offers several advantages. The system is relatively simple to operate and uses the least expensive reagent for SO_2 removal. The process does not produce any by-product and can tolerate significant fly ash in the system. The limestone scrubbing process is probably the safest to operate. However, there are disadvantages to this process. The system produces large quantities of solid waste that must be disposed of; sulfate scale forms in the scrubber and the demister; and corrosion and erosion affect the pumps, scrubber internals and reheater tubes. Also, the SO_2 removal efficiency of limestone scrubbers may be relatively low unless large amounts of limestone and very high liquid-to-gas ratios are used.

Lime flue gas scrubbers are gas-liquid contacting devices that use the chemical reactions between $Ca(OH)_2$ and SO_2 to remove sulfur oxides from the combustion gases generated during the operation of fossil-fired furnaces.

Lime scrubbers are advantageous because they are relatively simple to operate. Moreover, lime is more reactive than limestone, though it is also more expensive. The cost differential, however, also depends on the shipping distance, since lime weighs less than limestone on a molar basis. In

addition, the SO_2 removal efficiency of lime scrubbers is generally considerably higher than that of limestone systems on an equivalent cost basis. The lime scrubbing process can tolerate significant fly ash in the system. Disadvantages of the system include the production of large quantities of solid waste for disposal; the formation of sulfate scale in the scrubber and the demister; corrosion and erosion attack on the pumps, scrubber internals, and reheater tubes; and the safety problems of handling lime.

FOREWORD

This study was conducted for the U.S. Army Armament Materiel Readiness Command as a reimbursable project under Project No. M1-85262-M1-MP. The Technical Monitor was Mr. Thomas Wash.

The study was performed by the Environmental Divison (EN), U.S. Army Construction Engineering Research Laboratory (CERL). Dr. R. K. Jain is Chief of EN.

Appreciation is expressed to Dr. Harvey Rosenberg of Battelle's Columbus Laboratories, who was a source of technical information for this report. CERL participants in the project were Vinod Singh, Steve Struss, and Dr. Jim Stukel.

COL L. J. Circeo is Commander and Director of CERL, and Dr. L. R. Shaffer is Technical Director.

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POTENTIAL USE ON ARMY COAL-FIRED BOILERS

1 INTRODUCTION

Background

After passage of the Clean Air Act of 1970, strict limitations were
placed on the amount of SO_x emissions that could be released to the atmo-
sphere. Little proven technology was available at the time for SO_x removal
from stack gas at coal burning heating plants. This meant that coal burning
units either purchased low sulfur coal (which was then extremely hard to
find), or converted to a cleaner burning fuel such as gas or oil (which is now
extremely expensive). Many of the Army's boilers -- designed to fire coal but
for environmental reasons altered to burn gas and oil -- must now be converted
back to coal, with the additional cost of air pollution control equipment and
waste disposal.

The Army is returning to coal as a source of energy because of the fuel's
low cost, abundance, and domestic availability. Low sulfur Eastern coal is
used mainly for metallurgical purposes and is neither available nor economical
for steam generation. Low sulfur Western coal is abundant but transportation
costs and high ash content make it undesirable for use in the populous regions
of the United States. High-sulfur coal is abundant throughout most of the
United States and has a high energy content, but sulfur removal is a problem
that must be solved.

Coal preparation techniques commonly used today can remove some sulfur
and other impurities before the coal is burned, but these techniques are nei-
ther efficient nor proven for sulfur removal. Therefore, if high-sulfur coal
is burned, the flue gas must be cleaned to acceptable State and Federal stan-
dards before it can be discharged to the atmosphere.

Objective

The objectives of this study were to develop guidelines to help Army
installations select flue gas desulfurization (FGD) processes; to analyze and
evaluate existing FGD techniques which could be used in Army coal-burning
boilers (capacity range, 25 to 250 MBtu/hr [7.33 to 73.27 MW] heat input); and
to recommend systems that are most practical for Army installations.

Scope

This study was limited to six FGD systems which use processes applicable to Army-size industrial coal burning boilers: the limestone, lime, single-alkali, double-alkali, magnesium oxide, and citrate processes.

Approach

All existing FGD technologies were analyzed by an evaluation of their performance in the field, a literature search, and consultations with manufacturers. Through this assessment, CERL determined each technique's efficiency of SO_x removal, capital equipment cost, reliability, maintenance requirements, personnel requirements, secondary pollution potential, and adaptability to Army-size boilers. This information was then tabulated and extrapolated to reflect values for the 25 to 250 MBtu/hr (7.33 to 73.27 MW) boilers owned by the Army.

To validate the technical data, field evaluations were made on the six FGD systems applicable to Army-size boilers. When actual operational information was not available for a particular FGD system, estimates from manufacturers and vendors were used.

2 OVERVIEW OF AVAILABLE FGD SYSTEMS

A large number of organizations are working on the problem of removing SO_2 from stack gases. More than 50 individual processes have been developed in the United States (Figure 1);[1] however, these are of two basic types: regenerable processes recover SO_2 in some useful form of by-product, such as sulfuric acid or elemental sulfur; throwaway processes have as their end product a mixture of sulfur compounds and usually fly ash, intended for discharge as a waste material. Table 1 is a list of processes actively committed to full-scale FGD systems in the United States, and a similar list of scrubbing processes used on industrial boilers is given in Table 2.

Throwaway processes are at a more advanced stage of development than regenerable processes; still, a major problem is disposal of the sludge produced. And although throwaway processes are, on the whole, less complex and expensive than the regenerable processes, these cost differences disappear when the expense of indefinitely storing of sludge from the throwaway process is considered.

Regenerable FGD processes offer certain advantages over throwaway processes: (1) no sludge or filter cake for disposal; (2) SO_2, sulfur, or N_2SO_4 product; (3) significantly less secondary waste material. But these advantages are gained at the price of generally higher scrubber investment and scrubber operating costs, and increased energy input -- especially if a reducing gas is required for sulfur production. Also, more efficient particulate removal is required ahead of the SO_2 absorber to prevent fly ash from entering the regeneration loop. (In contrast, throwaway processes, with the exception of the double-alkali, can remove much or all of the fly ash with the SO_2.) In Table 3, leading regenerable FGD processes are defined by type (wet or dry), by-product produced, and reducing agent requirements. Dry processes have the advantage of eliminating the need for stack gas reheat.

FGD Throwaway Processes

There are four well-known and well-studied FGD throwaway processes: single alkali, wet limestone, wet lime, and double alkali. These processes are conspicuous because of their relative simplicity and their use of readily available raw materials. Except the single-alkali process, which produces a clear scrubbing liquor waste, all the processes have solid wastes as their end product.

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[1] Battelle's Columbus Laboratories, Status of Stack Gas Technology for SO_2 Control, Final Report, Part II to Electric Power Research Institute (EPRI), EPRI 209 (August 1975).
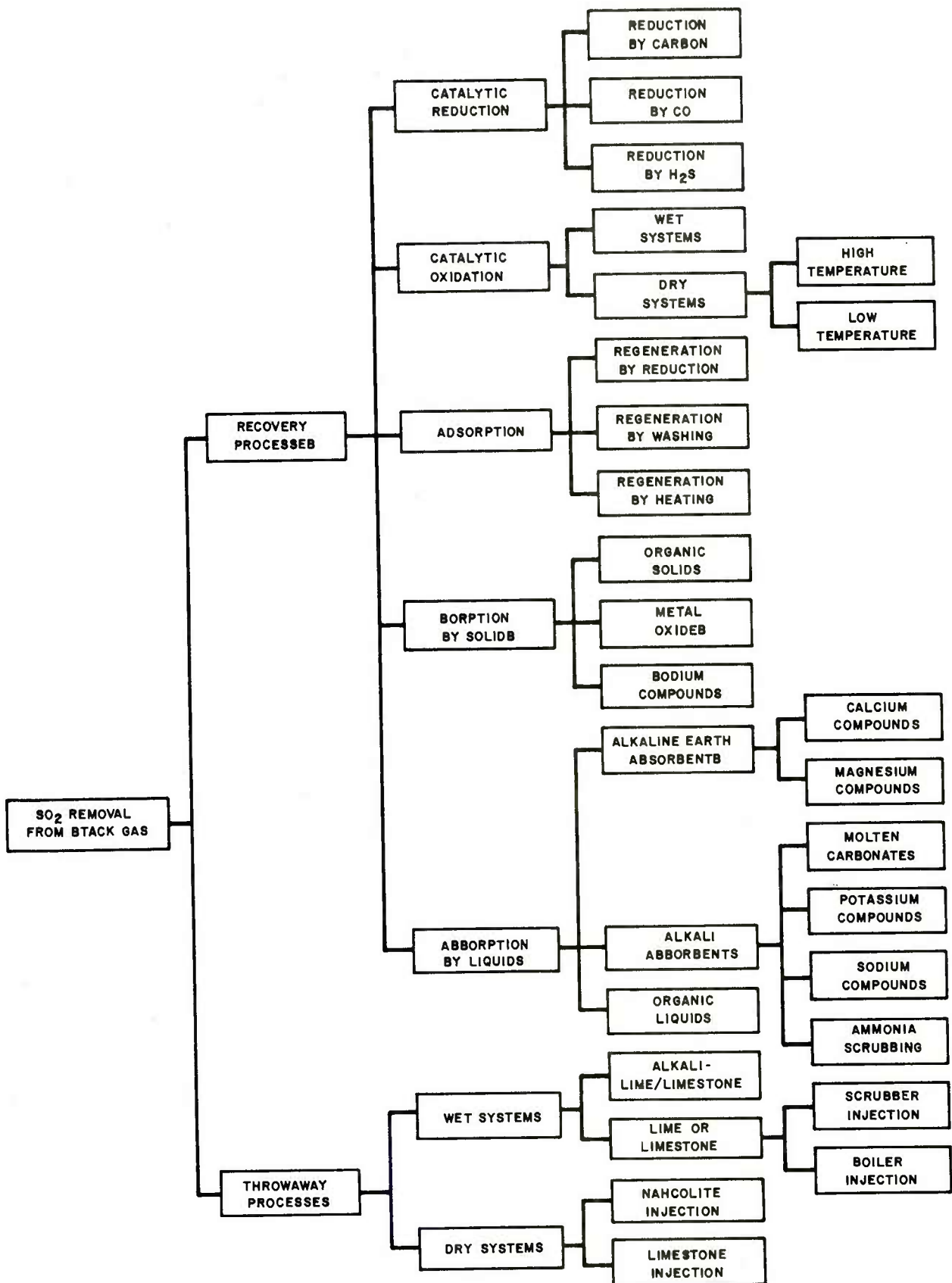
Figure 1. Technologies for the removal of SO_2 from stack gas.

Table 1

Chemical Process Types of Committed Active FGD Systems

(From B. A. Laseke, and T. W. Devitt, "Utility Flue Gas Desulfurization
in the U.S.," Chemical Engineering Process, Vol 76, No. 5
[May 1980]), p 48

| Process | FGD Capacity, MW | | | |
	Operational	Under Construction	Planned	Total
Limestone[a]	9,528	7,879	10,980	28,387
Lime[b,c]	7,082	5,633	4,266	16,981
Lime/Limestone[d]	20	708	600	1,328
Sodium Carbonate	375	550	0	925
Single Alkali	375	509	0	884
Magnesium Oxide	120	0	1,324	1,444
Wellman Lord	826	1,248	0	2,074
Dual Alkali	553	617	0	1,170
Aqueous Carbonate[e]	0	440	100	540
Citrate[f]	0	60	0	60
Total	18,879	17,644	17,270[g]	53,793

[a] Includes alkaline fly ash/limestone and limestone slurry process design configurations.
[b] Includes alkaline fly ash/lime and lime slurry process design configurations.
[c] Includes throwaway dry collection process design and throwaway wet scrubbing process design configurations.
[d] These systems are either prototypes designed to use lime or limestone, or full-scale systems in which the decision to use lime or limestone still has not been made.
[e] Includes throway dry saleable product (sulfur) dry collection process design configurations.
[f] This system is being installed at St. Joe Minerals' G.F. Wheaton Plant and is listed as a utility FGD system because the plant is connected by a 25-MW interchange to Duquesne Light Company.
[g] Because the processes for all planned systems are not known, the totals in this table are less than those in Table 1.

Table 2

Reported Industrial Use of FGD Equipment

Owner	Location	Process
Alyeska Pipeline Service Co.	Valdez, AK	Single alkali
American Thread	Marion, NC	Single alkali
Arco/Polymers	Monaca, PA	Conc. double alkali
Armco Steel	Middletown, OH	Lime
Canton Textiles	Canton, OH	Single alkali
Carborundum Abrasizes	Buffalo, NY	Lime
Caterpillar Tractor Co.	East Peoria, IL	Conc. double alkali
Caterpillar Tractor Co.	Joliet, IL	Dilute dbl. alkali
Caterpillar Tractor Co.	Mapleton, IL	Conc. double alkali
Caterpillar Tractor Co.	Morton, IL	Dilute dbl. alkali
Caterpillar Tractor Co.	Mossville, IL	Conc. double alkali
Chevron Oil Company	Bakersfield, CA	Single alkali
Firestone Tire and Rubber Co.	Pottstown, PA	Conc. double alkali
FMC (Soda Ash Plant)	Green River, WY	Single alkali
General Motors	Tonowanda, NY	Single alkali
General Motors	St. Louis, MO	Single alkali
General Motors	Parma, OH	Dilute dbl. alkali
General Motors	Pontiac, MI	Single alkali
General Motors	Dayton, OH	Single alkali
Georgia-Pacific Paper Co.	Crossett, AR	Single alkali
Great Southern Paper Co.	Cedar Springs, GA	Single alkali
Great Western Sugar	Billings, MT	Single alkali (NH_3)
Great Western Sugar	Findlay, OH	Single alkali
Great Western Sugar	Fort Morgan, CO	Single alkali (NH_3)
Great Western Sugar	Fremont, OH	Single alkali
Great Western Sugar	Gering, NE	Single alkali (NH_3)
Great Western Sugar	Greeley, CO	Single alkali (NH_3)
Great Western Sugar	Longmont, CO	Single alkali (NH_3)
Great Western Sugar	Loveland, CO	Single alkali (NH_3)
Great Western Sugar	Scotts Bluff, NE	Single alkali (NH_3)
Harris Mining Co.	Spruce Pine, NC	Single alkali
ITT Rayonier, Inc.	Fernandina Beach, FL	Single alkali
Kerr-McGee Chemical Corp.	Trona, CA	Single alkali
Mead Paperboard Co.	Stevenson, AL	Single alkali
Minn-Dak Farmer's Cooperative	Wahpeton, ND	Single alkali (NH_3)
Mobil Oil Company	San Ardo, CA	Single alkali
NCR-Appleton	Roaring Springs, PA	Single alkali
Nekoosa Papers, Inc.	Ashdown, AR	Single alkali
Quality Mills, Inc.	Mt. Airy, NE	
Rickenbacker Air Force Base	Columbus, OH	Lime/Limestone
Sheller Globe Corp	Norfolk, VA	Single alkali
St. Joe Minerals Corp.	Monaca, PA	Citrate
St. Regis Paper Co.	Cantonment, FL	Single alkali
Texaco Incorporated	San Ardo, CA	Single alkali
Texasgulf	Granger, WY	Single alkali
Transco Textiles, Inc.	Augusta, GA	Single alkali
U.S. Gypsum, Corp.	Oakmont, PA	Dilute dbl. alkali

Table 3

Selected Regenerable FGD Processes

Process	Type	By-Product	Reducing Agent	Current Status
Wellman-Lord	Wet	SO_2[1] H_2SO_4[1] Sulfur[1]	None None Reducing	In operation on a 115-MW coal-fired boiler natural gas is used to reduce SO_2
MgO	Wet	H_2SO_4	Coke[2]	In operation on 1/3 of the flue gas from a 315-MW coal-fired boiler.
Cat-Ox	Dry	80% H_2SO_4	None	Shut down after unsuccessful operation on a 110-MW coal-fired boiler.
Chiyoda	Wet	$CaSO_4 \cdot 2H_2O$[3]	None	Shut down after testing at 23-MW level on flue gas from a coal-fired boiler; in full-scale operation in Japan on oil-fired boilers
FW-BF[4]	Dry	Sulfur	Reducing gas or coal[5]	Shut down after unsuccessful operation at 20-MW level on flue gas from a coal-fired boiler; successfully operated in Germany on coal-fired boiler at 30-MW level.
Shell	Dry	SO_2[1] H_2SO_4[1] Sulfur[1]	H_2[6] H_2[6] Reducing gas and/or H_2[6]	Pilot plant in operation on a 0.6 MW slipstream from coal-fired boiler; full-scale operation at oil refinery in Japan.
Catalytic IFP[7]	Wet	Sulfur	Reducing gas	Operated in France at 30-MW level on flue gas from oil-fired boilers.
Citrate	Wet	Sulfur	H_2S	100-MW demonstration unit under construction on coal-fired industrial boiler.
Al-ACP[8]	Wet	Sulfur	Coke	10 to 100-MW demonstration unit will be constructed on coal-fired utility boiler.

(1) FGD process concentrates SO_2 for further processing to H_2SO_4 or elemental sulfur.
(2) Coke is needed for $MgSO_4$ decomposition.
(3) Gypsum is probably not a marketable by-product in the U.S.
(4) Foster-Wheeler/Bergbau Forschung.
(5) RESOX process developed by Foster Wheeler uses coal to reduce SO_2 to sulfur.
(6) Hydrogen is required to regenerate the SO_2 acceptor.
(7) Catalytic, Inc.-Institut Francais du Petrole.
(8) Atomics International-Aqueous Carbonate Process.

Single alkali, a clear liquor scrubbing process, principally used on relatively small industrial boilers, eliminates the problem of scrubber scaling, permits the use of more efficient scrubbers operating at smaller liquid-to-gas ratios, and generally has a higher efficiency for SO_2 removal than the other systems, which use slurry scrubbing processes. On the other hand, the slurry processes are used more on large electric power-plants and, in general, are more complex than the single-alkali process. The wet scrubbing processes (single alkali, lime, and limestone) have one disadvantage in common: stack gas reheat may be needed to increase plume buoyancy. Dry sorption processes avoid this problem because no stack gas reheat is required. Wet scrubbing processes may or may not need reheat if the initial stack gas temperature is high enough to evaporate the water in the gas and still maintain plume buoyancy. Depending on the process requirements, it may be possible to bypass a portion of the flue gas around the FGD equipment and use this portion of gas to reheat scrubbed gas, thus maintaining plume buoyancy. Of the 22 FGD systems reported on industrial boilers in the United States at the end of 1977, 17 were single-alkali processes, 4 were double-alkali processes, and 1 employed lime scrubbing.[2]

Single Alkali

The single-alkali process is the simplest of the four, and offers high SO_2 removal efficiency at low liquid-to-gas ratios in the scrubber. However, it has major disadvantages: NA_2CO_3, NaOH, or NH_4OH are expensive alkali sources, and water soluble salts produced in the process must be disposed of in an environmentally acceptable manner. Consequently, the process has limited applicability and can be used only in arid regions or where local governments permit the discharging of soluble salts. The latter case, of course, is unlikely; thus the process is limited mainly to the deserts of the western United States.

Wet Limestone and Lime Scrubbing

The most notable advantages of the wet limestone and lime scrubbing processes are their use of a relatively inexpensive source of alkali and their relative simplicity. In the past, the limestone process has been plagued by operational problems, mainly scaling in the scrubber and plugging of the mist eliminator. In the lime scrubbing process, the major success for high-sulfur coal has occurred using carbide lime -- a special reagent and generally unavailable. Thus, carbide lime scrubbing can be used only when an installation has an assured supply of reactant. However, this would no longer be a problem if hydrated lime were shown to produce results similar to carbide lime on a full-scale, high-sulfur, coal-fired, closed-loop system.

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[2] H. S. Rosenberg, "How Good is Flue Gas Desulfurization?", Hydrocarbon Processing (May 1978), pp 132-135. (Adapted from a paper presented at the 43rd Midyear Refining Meeting, American Petroleum Institute, Toronto, Canada, May 8-11, 1978).

Double Alkali

The double alkali process attempts to avoid scrubber scaling and mist eliminator plugging by putting in a second liquor loop with soluble alkali (either Na+ or NH_4+ ion) as the scrubbing liquor. The scrubbing liquor is then neutralized with lime or limestone outside the scrubbing circuit, and every effort is made to keep Ca++ ion out of the scrubbing circuit. In the double-alkali scrubbing process, the basic chemistry is quite complex. The latest version incorporates a third loop using an ion exchange resin to ensure no scaling in the scrubber.[3]

Regenerable FGD Processes

MgO Scrubbing

The greatest advantage of the MgO scrubbing process is that there are no major waste streams to be disposed of, and the H_2SO_4 produced may be of market quality. The SO_2 is absorbed by the MgO slurry, the sulfited slurry is centrifuged, the magnesium sulfite dried, and then calcined to reform the MgO and liberate the SO_2. The SO_2 is used to produce H_2SO_4 in an acid plant, and the MgO is recycled to the power plant.

Since the acid is of high technical quality, part of the operating costs can be recovered by selling the acid. However, the user of the process must have a market for H_2SO_4 since storage costs are prohibitive. This greatly diminishes its effectiveness and market potential. Also, it is very difficult to separate the fly ash from the recovery liquor because of the presence of insoluble magnesium salts. If the fly-ash collectors malfunction, the process would probably have to be shut down.

Catalytic Oxidation

Like MgO scrubbing, the catalytic oxidation process produces H_2SO_4, and the process user must have a market for the product. Basically, the process involves heating the flue gas to 850°F (454°C) and passing it over a catalyst in the presence of oxygen, which converts the SO_2 to SO_3; then the SO_3 is absorbed in acid scrubbers. Unfortunately, the acid may be of low quality, and this makes it difficult to sell. Furthermore, since the catalyst is readily deactivated, fly ash must be kept out of it.

The catalytic oxidation process is not easily retrofitted because the operating temperature required for oxidation of SO_2 to SO_3 requires complicated stack gas reheat methods. The advantages of the process are that it

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[3] R. E. Anderson, *The Triple Alkali System for Removal of Sulfur Oxides from Stack Gases Using Ion Exchange* (Diamond Shamrock Chemical Company, 1974).

uses relatively simple equipment, requires no raw material, and produces no major stream of solid waste.

Wellman-Lord

The Wellman-Lord process involves absorbing the SO_2 in a solution of soluble sodium bisulfite to form sodium sulfite. By heating this salt (which is also soluble) in a separate vessel, the absorbed SO_2 can then be thermally stripped from the solution. The SO_2 can then be recovered or converted to other sulfur chemicals. The Wellman-Lord process can therefore also produce either sulfuric acid or elemental sulfur (the more environmentally acceptable by-product), which must be sold or disposed of.

In the Wellman-Lord process in which sulfur is recovered, a reducing agent is required to convert stripped SO_2 to elemental sulfur. The most convenient reducing agent, and the one selected for a Northern Indiana Public Service Company (NIPSCO) demonstration plant examined by CERL, is natural gas. Natural gas, however, is in short supply, quite expensive, and not well suited for power plant use. To permit regeneration of the sodium sulfite scrubbing liquor, fly ash must be kept out of the regeneration system. This requires a fly-ash collection device that is extremely efficient and reliable. The process also requires a purge stream to remove sulfate from the system and has rather high steam requirements due to evaporation of water in the evaporator/crystallizer.

Chiyoda

The Chiyoda process originally used dilute sulfuric acid as the scrubbing liquor and then neutralized the acid external to the scrubber with limestone. The process thus tried to prevent scrubber scaling and mist eliminator plugging by eliminating formation of Ca++ ions in the scrubber circuit. The end product is dry gypsum, which is marketed in Japan; but due to the uncertainty of this market in the United States, the material is usually disposed of as a waste. The greatest disadvantage of the process is that it scrubs an acid gas with an H_2SO_4 solution. Thus, very high liquid-to-gas ratios must be used; and this leads to high power requirements, very large scrubbers and oxidizer towers, and, in general, a large volume scrubber circuit. Also, large quantities of acid must be handled and acid-resistant construction materials are required. A modification of the process is now being developed to correct these shortcomings.

Foster-Wheeler, Bergbau-Forschung

The Foster-Wheeler, Bergbau-Forschung (FW-BF) process involves dry adsorption of SO_2 on char followed by thermal regeneration to produce a concentrated stream of SO_2. The loaded sorbent is conveyed to the desorber where hot sand in an inert atmosphere heats the carbon to 1100^0F (593^0C) to release SO_2. The sand and carbon are then separated on a vibrating screen, and the sand is recycled through a heater. The desorbed carbon is cooled and returned

to the top of the absorber. BF claims that the process works well with fly ash in the flue gas. Other advantages claimed include good flexibility in design, NO_x removal capability (40 to 60 percent by adsorption on char), and no stack gas reheat.

The FW-BF process produces a concentrated SO_2 stream which can be used to produce sulfuric acid or elemental sulfur. Foster-Wheeler has developed what they call the RESOX[TM] process, which has a fixed bed of anthracite coal to reduce the SO_2 to sulfur. This process has been used on a 20-MW coal-fired boiler at Scholz and on a 30-MW coal-fired boiler in Germany, where about 100,000 scfm (47.19 m^3/sec) of flue gas, containing 1000 parts per million (ppm) SO_2, is treated with 80 percent removal efficiency.

Shell (Copper Oxide)

In the dry, high temperature Shell (copper oxide) process, the SO_2 in flue gas is concentrated by reacting it with CuO to form CuSO4, and then regenerating (reducing) the sulfate with hydrogen to yield CuO and SO_2. The concentrated SO2 stream can be used to produce sulfuric acid or elemental sulfur. The CuO is supported on small alumina pellets which are held by wire screens in a specially designed fixed-bed reactor. As the flue gas passes through the acceptance reactors, the SO_2 breaks through in a typical plug flow fashion. (The cycle length is determined by the desired extent of SO_2 removal.) The flue gas is then switched to another reactor, and the loaded acceptor is regenerated with a mixture of hydrogen and steam.

The major drawback to this process is the hydrogen needed to regenerate the sorbent. The hydrogen can be produced by steam reforming, partial oxidation, or gasification of a suitable feedstock, but this and the related gas cleanup operations will add greatly to the complexity and cost of the overall process.

Most of the experience with the Shell process has been on flue gas from oil-fired boilers, although some units have been operated on slipstreams from coal-fired boilers in the Netherlands and in Florida (Tampa Electric).

Fumeless Ammonia

The Catalytic, Inc., and Institut Francais du Petrole process (Catalytic-IFP) or fumeless ammonia scrubbing process, is an attempt to avoid the plume problem experienced when scrubbing SO2 with an aqueous ammonia solution. The very visible plume results from ammonia-related species in the scrubbed gas. This process uses an ammonia-based system to scrub SO_2; and then, with the IFP reducing process, regenerates the spent ammonium salts and produces sulfur. Ammonia concentration, pH, and liquid-to-gas ratio are controlled to eliminate the "blue" plume from the scrubbing system. SO_2 and NH_3 are recovered from ammonium sulfite and bisulfite by evaporation, and from ammonium sulfate by reduction with sulfur. Carbon monoxide and hydrogen are also added to produce the desired concentration of H_2S for the Claus plant,

where elemental sulfur will be generated. The process is reported to have a greater than 90 percent SO_2 removal efficiency; however, it has not been tested beyond the pilot plant scale (an equivalent of 30-MW flue gas from two 250-MW oil-fired boilers at the Champagne sur Oise Power Plant near Paris).

Aqueous Carbonate

The Atomic International's aqueous carbonate process (AI-ACP) uses a solution of Na_2CO_3 as the scrubbing medium. The key component of the ACP scrubbing system is a modified spray dryer which serves as a reaction chamber for the removal of SO_2. In the spray dryer, sodium carbonate solution is atomized by one or more high-speed centrifugal atomizers, and mixed with the hot gas entering the dryer through vane-rings which surround the atomizers. The fine mist of solution droplets absorbs SO_2 while at the same time the thermal energy of the waste gas vaporizes the water without saturating or excessively cooling the gas. The reaction product is a dry powder containing Na_2SO_3 and Na_2SO_4. This product is reduced with coke (or other carbon source) in a molten salt reactor. The molten product from this reactor is quenched and dissolved in water, and the resulting solution is cooled, filtered, and reacted with CO_2 (from the reduction reactor) to liberate H_2S. Then, the H_2S-rich gases are sent to a Claus plant for recovery of elemental sulfur. The carbonation generates sodium carbonate for recycle to the scrubber. Although both the scrubbing system and the reduction reactor have been tested separately, the entire process has not yet been demonstrated in a single installation. Development efforts to date have concentrated only on pilot-plant scale.

Citrate

The citrate process involves selective absorption of SO_2 in a citric acid solution followed by conversion of the absorbed sulfur into elemental sulfur. The flue gas is scrubbed with an aqueous solution of sodium citrate, citric acid, and sodium thiosulfate. These compounds buffer the solution, thereby increasing the absorption capacity, and inhibit oxidation of the sulfite to sulfate. The resulting solution, which contains about 10 g SO_2/L, is contacted with H_2S in a stirred reactor, thereby reducing the SO_2 to sulfur. The sulfur is separated by oil or froth flotation and melted, and the citrate solution is recycled to the absorber. The H_2S is normally generated by reducing sulfur, in which case two-thirds of the total sulfur produced is recycled in this manner.

The process was pioneered by the U.S. Bureau of Mines (USBM) Salt Lake City Metallurgy Research Center. The USBM pilot plant at Bunker Hill lead smelter, Kellog, ID, is the only pilot plant operating now. This plant uses an H_2S generator with natural gas as the feedstock.

With the exception of the citrate magnesia processes, these regenerative processes are believed to be either too complicated, not sufficiently tested,

or unsuitable for retrofit operation, and will not be discussed further in this report. Some organizations in the United States which can supply such technology are listed in Table 4.

Table 4

Typical Suppliers of FGD Equipment

1. Air Pollution Industries
    (Now a subsidiary of
    Neptune International)

2. Air Products
    P.O. Box 538
    Allentown, PA 18105

3. Babcock & Wilcox
    P.O. 2423
    North Canton, OH 44720

4. Carborundum Environmental Systems
    Niagara Falls, NY 14302

5. The Ceilcote Company
    140 Sheldon Road
    Berea, OH 44017

6. Combustion Equipment
      Associates, Inc.
    555 Madison Avenue
    New York, NY 10022

7. Combustion Engineering, Inc.
    1000 Prospect Hill Road
    Windsor, CT 06095

8. Davy Powergas, Inc.
    P.O. Box 36444
    Houston, TX 77036

9. Dresser Industries
    P.O. Box 19566
    Irvine, CA 92713

10. Ducon Company, Inc.
    147 East Second Street
    Mineola, NY 11501

11. Entoleter, Inc.
    P.O. Box 1919
    New Haven, CT 06509

12. Envirotech Corporation
    One Penn Plaza
    New York, NY 10001

13. FMC Environmental Equipment
    Box 276
    Hudson, OH 44236

14. Heater Technology, Inc.
    P.O. Box 1420
    Ponca City, OK 74601

15. Joy Industrial Equipment Co.
    4565 Colorado Boulevard
    Los Angeles, CA 90039

16. Koch Engineering
      Company, Inc.
    161 East 42nd Street
    New York, NY 10017

17. Neptune International Corp.
    145 Cedar Lane
    Englewood, NJ 07631

18. Arthur G. McKee and Company
    10 South Riverside Plaza
    Chicago, IL 60606

19. Morrison-Knudsen Co., Inc.
    P.O. Box 7808
    Boise, ID 83729

20. Peabody Engineered Sys-
      tems, Inc.
    835 Hope Street
    Stamford, CT 06907

21. Riley Environeering, Inc.
    4233 North United Parkway
    Schiller Park, IL 60176

22. Research-Cottrell
    P.O. Box 1500
    Somerville, NJ 08876

23. Rockwell International
    8900 DeSoto Avenue
    Canoga Park, CA 91304

Table 4 (Cont'd)

24. W. W. Sly Manufacturing Co.
    P.O. Box 5939
    Cleveland, OH 44101

25. Swemco, Inc.
    470 Park Ave. South
    New York, NY 10016

26. United Engineers and
    Constructors, Inc.
    30 South 17th Street
    Philadelphia, PA 19101

27. Universal Oil Products
    P.O. Box 1107
    Darien, CT 06820

28. Whellabrator Frye, Inc.
    600 Grant Street
    Pittsburgh, PA 15219

29. Zurn Industries, Inc.
    P.O. Box 2206
    Birmingham, AL 35201

This chapter describes in detail some of the characteristics of the six processes which may be considered for use at military installations (Table 5). It should be noted that some of these processes can be operated as "wet" or "dry" scrubbers -- i.e., they can be designed to produce a dry or wet sulfur-containing by-product, although all currently operating systems at either industrial or steam-electric plants are based on conventional wet scrubbing.

Single-Alkali Scrubbers

Definition

Single-alkali flue gas scrubbers are gas-liquid contacting devices that use the chemical reactions between soluble alkali -- such as $NaOH$ or Na_2CO_3 -- and SO_2 to remove the oxides of sulfur from the combustion gases generated during the operation of fossil-fired furnaces.

Application Areas

Single-alkali scrubbers can clean flue gases from either coal-fired or oil-fired boilers. Though sometimes scrubbers are also expected to remove particulates, the main function of a single-alkali scrubber is to ensure that only a minor amount of SO_2 is allowed to escape with the flue gas. The level of allowed SO_2 emission varies with the local air pollution control regulations. Single-alkali scrubbing is now the most widely employed process on industrial boilers, but the process produces a significant liquid waste by-product which may be a disposal problem.

Principle of Operation

A simplified flow diagram of a typical single-alkali scrubbing system is shown in Figure 2. Flue gas from the boiler, after passing through a suitable collector for particulate removal, enters the lower part of the scrubber. The flue gas then passes up to a "reaction zone" (which could be a venturi, or a system of trays and packings, a spray zone, or a combination of these), where the gas is mixed vigorously with the alkali solution showering down. The gas attains its adiabatic saturation temperature, while the SO_2 is removed according to the overall chemical reaction:

$$2NaOH + SO_2 + 1/2\ O_2 \rightarrow Na_2SO_4 + H_2O$$
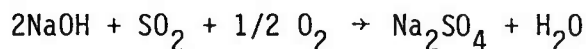
Intermediate products include sodium sulfite and bisulfite.

After this, the flue gas passes through a mist eliminator zone, where the entrained liquids and dissolved salts are captured, collected, and washed back to the scrubbing section. The essentially droplet-free scrubbed gas may then

Table 5

Chemical Process Types at United States Industrial Locations
(From John Tuttle and R. Michael McAdams, "The Status of
Industrial Boiler Applications in the United States," paper
presented at the USEPA Las Vegas FGD Symposium, n.d.).

| Process | FGD Capacity, 10^3 scfm (m^3/sec) | | | |
	Operational	Construction	Planned	Total
Single Alkali	4,770 (2.251)	656 (0.310)	286 (0.135)	5,712 (2.700)
Limestone	55 (0.026)	0 (0.000)	320 (0.151)	375 (0.177)
Lime	84 (0.040)	0 (0.000)	30 (0.014)	114 (0.054)
Double Alkali	591 (0.279)	437 (0.206)	0 (0.000)	1,028 (0.485)
Citrate	0 (0.000)	142 (0.067)	0 (0.000)	142 (0.067)
Other/Unspecified	0 (0.000)	0 (0.000)	1,334 (0.630)	1,334 (0.630)
	5,500 (2.596)	1,235 (0.583)	1,970 (0.930)	8,705 (4.108)

27
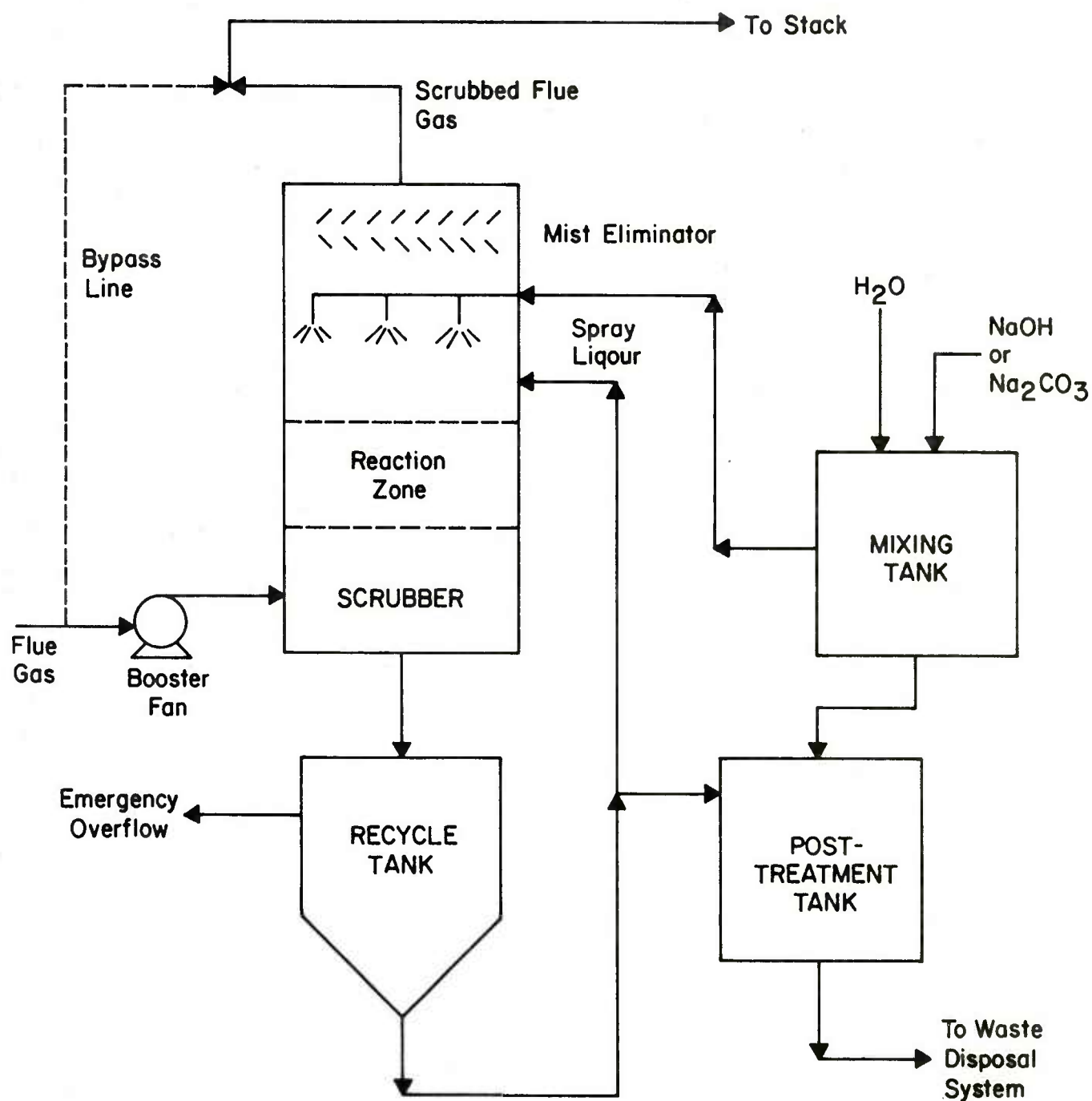
Figure 2.   Single-alkali scrubbing system.

be reheated, for both buoyancy and stack lining corrosive protection, before venting through a stack and into the atmosphere.

The effluent from the scrubber bottom enters a recycle tank. Some of this liquid can be returned to the scrubbing system; the rest is pumped as waste to the post-treatment tank, where the sodium sulfite and bisulfite is converted to sulfate by reacting with air. The post-neutralization tank contents are then discharged to the waste disposal system.

Advantages and Disadvantages

Advantages of the single-alkali scrubbing process include its ease of operation, its highly reactive agent, and its freedom from plugging and scaling. The process also does not produce any solid by-product and can tolerate some fly ash in the system. However, the process has some disadvantages -- especially waste disposal. Since sodium salts are soluble, the liquid waste must usually be treated before it can be discharged to the environment. The cost of reagents used, whether $NaOH$ or Na_2CO_3, is also frequently much higher than SO_2 scrubbing reagents such as lime and limestone. The use of alkali, especially caustic ($NaOH$), also requires increased concern with safety procedures for operating personnel.

Efficiency

The SO_2 removal efficiency of single-alkali scrubbing systems is generally high, in the range of about 85 percent to 99 percent. Major factors affecting this efficiency include the coal's sulfur content, the stoichiometry used, number of scrubbing stages, and the liquid-to-gas ratio. Relatively low liquid-to-gas ratios can be used because of the solubility and reactivity of the chemicals involved. In a venturi-contacting device, the SO_2 transfer between gas and liquid phases is primarily a function of pressure drop over a wide range of liquid-to-gas ratios and SO_2 contents, as long as excess soluble alkali is present in the waste liquor. Figure 3 shows SO_2 removal as a function of venturi pressure drop when sodium carbonate solution is used as the reactant. These data were obtained in a 100×10^6 Btu/hr (29.32 MW) equivalent system. Similar behavior is exhibited by other single-alkali scrubbing solutions when a venturi contact device is employed.

Reliability

Since the single-alkali scrubber is relatively free from plugging and scaling problems, it is highly reliable. In most units currently operated, the scrubber availability has been rated at above 90 percent.

Operational History

No figures on functional operating life have been recorded for the single-alkali scrubber. The first units in the United States went on stream
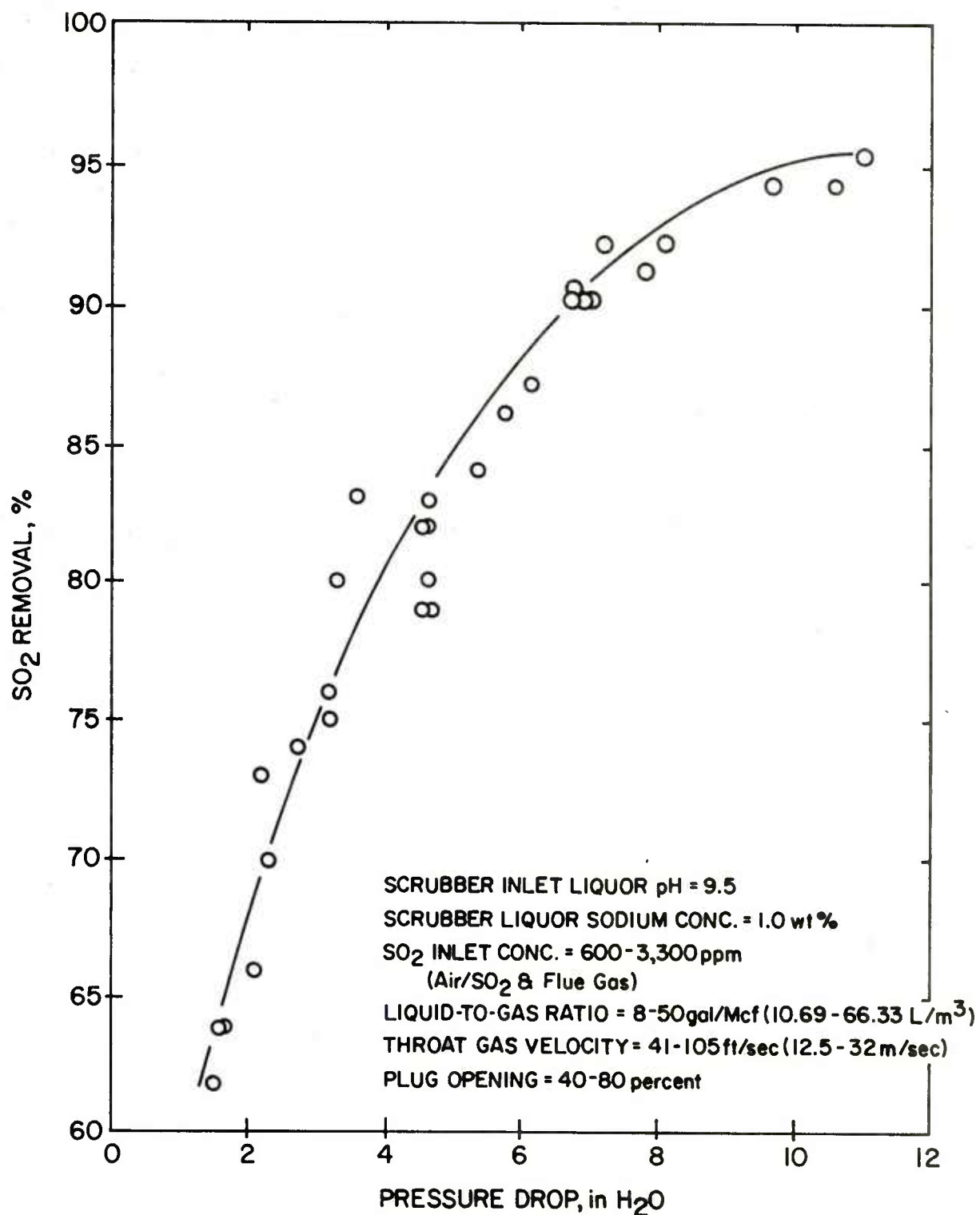
Figure 3. Effect of pressure drop on SO_2 removal efficiency in a single-alkali venturi scrubber. (From EPA Alkali Scrubbing Test Facility: Summary of Testing Through October 1974, EPA Report 650/2-75-047 [EPA, June 1975].)

in about 1972. The process is the easiest to control of all scrubbing systems.

Energy Impact

Single-alkali scrubbers require little energy to operate (averaging about 2 to 4 percent of the plant energy input). The pressure drop of the flue gas through the system will generally not exceed 15 to 20 in. (3733 to 4977 Pa) of water, which is not excessive, but generally does require some added blower or fan capacity at the heating plant.

Factors Affecting Performance

The oxygen content of the flue gas will directly affect the amount of sulfate produced. In some units, additives such as manganese and other heavy metal salts have been included as an oxidation catalyst in the feed to the post-neutralization tank. The major problem with this process is that the effluent, Na_2SO_4, must eventually be discharged somewhere. The only practical means of disposal is by standard municipal sewage treatment or deep well injection. Either method requires advance planning and technical verification that the disposal method is nonpolluting. Elsewhere, disposal in municipal sewers or in deep wells may be possible.

Retrofit Considerations

Single-alkali scrubbing processes are easy to retrofit because less auxiliary equipment is required than for scrubbing processes which produce a solid waste product. Since bypass capability is often desirable, the scrubbing system can be installed parallel to the ductwork, anywhere between the mechanical collector and the stack. In some systems the scrubber has its own stack, and flue gas from more than one boiler can be combined in a manifold duct leading to a scrubbing system, separate from the plant. Coarse particulate removal and added fan capacity must normally be included in retrofit plans.

Cost Profile

Table 6 presents the estimated ranges of capital and recurring cost elements for single-alkali scrubbers. Capital costs are installation expenses in FY78 dollars for a single-alkali scrubbing system consisting of a venturi and tray separator, with an associated reagent handling system and scrubbing liquor circuit. The costs, however, exclude any ponding or flue gas reheat system. The capital cost retrofit multiplier is at the bottom of the table. The operating costs are based on a capacity factor of 80 percent (7008 hours/year) and include recommended operating labor, maintenance, electricity, process water, soda ash, and waste disposal requirements.

Table 6

Estimated Costs of New Single-Alkali Scrubbing Systems
(Coal of 0.4-4.5 Percent Sulfur)

Boiler Input Heat Rate, 10^6 Btu/hr (MW)	Capital Cost ($ Million)*	Operational Cost, $/ton Coal Burned ($/MT)**	Soda Ash, K ton/yr (kt/yr)
25 (7.325)	0.5-0.8	12-33 (13.2-36.3)	0.08- 1.0 (0.07-0.91)
50 (14.65)	0.8-1.2	8-26 (8.8-28.7)	0.16- 2.0 (0.15-1.81)
100 (29.30)	1.2-2.0	5-23 (5.5-25.4)	0.30- 4.0 (0.27-3.63)
150 (43.95)	1.7-2.7	5-21 (5.5-23.1)	0.50- 6.0 (0.45-5.44)
250 (73.25)	2.5-3.9	4-20 (4.4-22.0)	0.80-10.0 (0.73-9.1)

*Marshall and Swift Cost Index = 561.8 (October 1978).

**No capital-related costs, such as steam for flue gas reheat, are considered.

Note: Retrofit factor = 1.00 to 1.80, dependent on boiler size and age. The smaller and older the boiler, the closer this factor approaches 1.80. This factor should be used to multiply the above capital costs for new plant when retrofitting a single-alkali scrubber to an existing plant.

Limestone Scrubbers

Definition

Limestone flue gas scrubbers are gas-liquid contacting devices that use the chemical reactions between limestone (mostly $CaCO_3$) and SO_2 to remove the oxides of sulfur from the combustion gases generated during the operation of fossil-fired furnaces.

Application Areas

Limestone scrubbers can be used to clean flue gases from either coal-fired or oil-fired boilers. Though sometimes scrubbers are also expected to remove particulates, the main function of a limestone scrubber is to ensure that most of the SO_2 is removed before flue gas leaves the stack. The level of allowed SO_2 emission varies with the local air pollution control regulations. Limestone scrubbers are used when incomplete SO_2 removal can be tolerated but significant liquid discharge streams are prohibited.

Principle of Operation

A simplified flow diagram of a typical limestone scrubbing system is shown in Figure 4. Flue gas from the boiler, after passing through a suitable collector for particulate removal, enters the lower part of the scrubber. The flue gas then passes up to a "reacting zone" (which could be a venturi, or a system of trays and packings, a spray zone, or a combination of these), where the gas is mixed vigorously with the limestone slurry showering down. The gas attains its adiabatic saturation temperature, while the SO_2 is removed according to the overall chemical reaction:

$$CaCO_3 + SO_2 + 1/2 \ H_2O \rightarrow CaSO_3 \cdot 1/2 \ H_2O + Co_2$$

Some calcium sulfate is also formed, and with forced oxidation, gypsum becomes the primary product.

After this, the flue gas passes through a mist eliminator zone, where the entrained liquids and solids are captured, collected, and washed back to the scrubbing section. The essentially droplet-free scrubbed gas may then be reheated, for both buoyancy and stack lining corrosive protection, before venting through a stack into the atmosphere.

The spent slurry, which is a mixture of unused limestone, fly ash, calcium sulfite, calcium sulfate, and sulfate solution, is pumped out at the scrubber bottom -- usually to a settling device (e.g., a settling or thickener tank). The overflow from this separation tank is normally recycled and mixed with the freshly prepared limestone slurry in a delay or reaction tank. The underflow from the system is disposed of as sludge (35 to 50 percent solids). A vacuum filtration system is also sometimes used with limestone scrubbers.
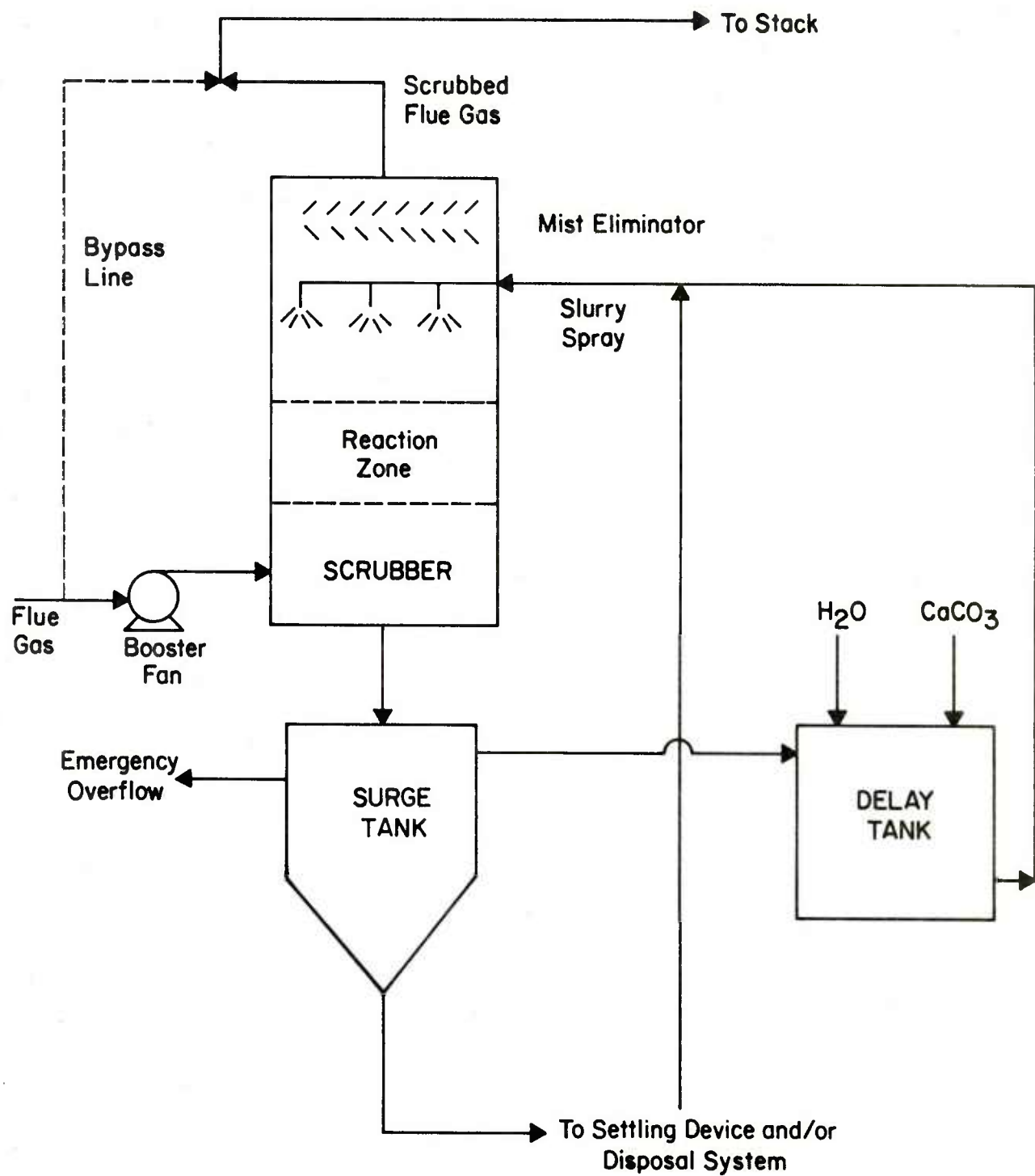
Figure 4. Limestone scrubbing system.

Advantages and Disadvantages

Wet limestone scrubbing offers several advantages. The system is relatively simple to operate and uses the least expensive reagent for SO_2 removal. The process does not produce any by-product and can tolerate significant fly ash in the system. The limestone scrubbing process is probably the safest to operate. Disadvantages include the additional expenditures necessary to purchase and maintain handling and crushing equipment. The system produces large quantities of solid waste that must be disposed of; sulfate scale forms in the scrubber and the demister; and corrosion and erosion affect the pumps, scrubber internals, and reheater tubes. Also, the SO_2 removal efficiency of limestone scrubbers may be relatively low unless large amounts of limestone and very high liquid-to-gas ratios are used.

Efficiency

The SO_2 removal efficiency of limestone scrubbers has been reported in the range of about 70 to 95 percent, though the higher end of the range is not normally achieved without staged scrubbing. Major factors affecting this efficiency include the coal's sulfur content, the stoichiometric ratio of limestone to SO_2 used, and the liquid-to-gas ratio.

These interrelationships are shown in Figure 5 for a 2 to 3 percent sulfur coal. A stoichiometry of about one (Ca/S ratio equal to 1.0) results in SO_2 removals of about 60 to 80 percent in a two-stage contactor. Increasing liquid-to-gas ratio increases the SO_2 removal and limestone utilization at a given stoichiometric ratio.

Reliability

In general, the reliability of limestone scrubbers is still a problem. In closed-loop operation, plugging and scaling difficulties have been responsible for most limestone scrubber outages -- ranging anywhere from 10 to 50 percent downtime annually. Extra operating skill and maintenance are usually required during the winter to keep the limestone scrubber on-line. If an open-loop operation can be designed into the combined scrubber and waste disposal system, reliability can be considerably improved.

Operational History

No figures on functional life have been recorded for the wet limestone scrubber; most operating units are still in the demonstration period. The first units in the United States went on stream in about 1972. Generally, though, the higher the level of sulfur in the coal, the more difficult it is to achieve regulation compliance using this method of control. $CaCO_3$ is not highly reactive in sulfur removal, and as SO_x concentration goes up, removal efficiency decreases.
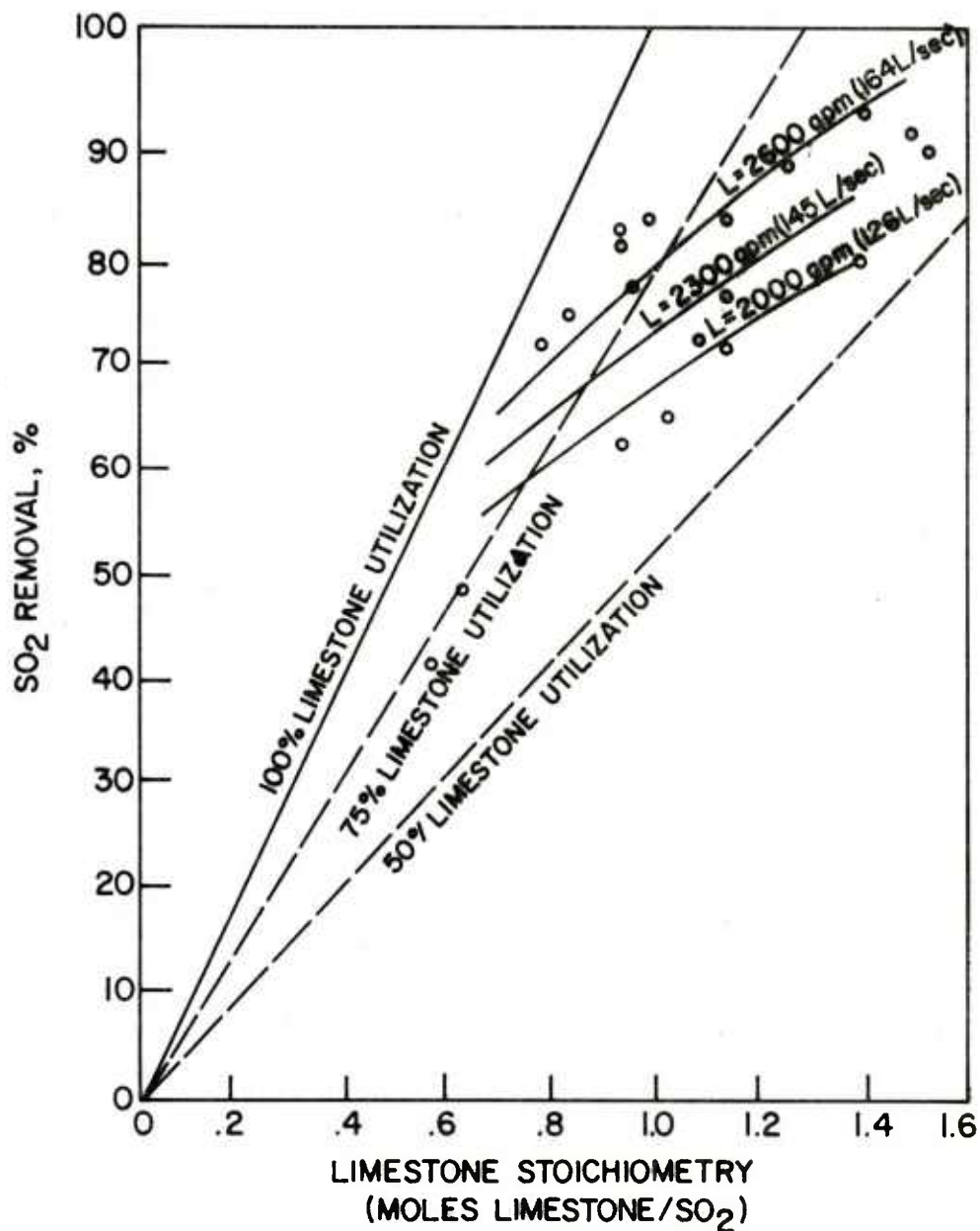
Figure 5. SO_2 removal efficiency as a function of limestone/SO_2 stoichiometry and slurry pumping rate in a two-stage contactor (From E. L. Biedell, R. J. Ferb, G. W. Malamud, C. D. Ruff, and N. J. Stevens, EPA Evaluation of Bahco Industrial Boiler Scrubber System at Rickenbacker AFB, EPA Report 600/7-78-115 [EPA, June 1978]).

Energy Impact

Limestone scrubbers require little energy to operate (average about 3 to 7 percent of the plant energy input). The pressure drop of the flue gas through the system will generally not exceed 20 in. (4977 Pa) of water, which is not excessive, but generally does require some added blower or fan capacity at the heating plant.

Factors Affecting Performance

Plugging and scaling are the major problems encountered in limestone scrubbers. Though plugging may be prevented by increasing the washing water flow, and scaling can be reduced by adding an oxidation inhibitor (e.g., MgO) to the scrubber slurry, these corrective measures may not be acceptable because they can cause water pollution when scrubber wastes are disposed of. Increasing the limestone/SO_2 stoichiometric ratio will generally improve the scrubber performance, but may not be the most economical approach to environmental control.

Retrofit Consideration

Limestone processes are relatively easy to retrofit because they require much less auxiliary equipment than do regenerative processes. Since bypass capability is often desirable, the scrubbing system can be installed parallel to the duct work, anywhere between the mechanical collector and the stack. In systems where the scrubber has its own stack, flue gas from more than one boiler can be combined into a manifold duct leading to a scrubbing system separate from the plant. Coarse particulate removal and added fan capacity must normally be included in retrofit plans. Space for waste disposal is generally a major consideration for limestone systems.

Cost Profile

Table 7 presents the estimated ranges of capital and recurring cost elements for limestone scrubbers. Capital costs are installation costs in FY78 for a limestone scrubbing system consisting of a venturi and a spray tower with associated limestone-handling system and slurry circuit. The costs, however, exclude the sludge pond and flue gas reheat system. The capital cost retrofit multiplier is shown at the bottom of the table. The operating costs shown are based on a capacity factor of 80 percent (7008 hours/year) and include recommended operating labor, maintenance, electricity, process water, limestone, and waste disposal requirements. Capital-related charges and steam reheat costs are not incorporated in this tabulation.

Table 7

Estimated Costs of New Limestone Scrubbing Systems
(Coal of 0.4-4.5 Percent Sulfur)

Boiler Input Heat Rate, 10^6 Btu/hr (GJ/hr)	Capital Cost ($ Million)**	Operational Cost, $/ton Coal Burned ($/MT)**	Limestone, K ton/yr (kt/yr)
25 (26.4)	0.6-1.1	14-24 (15.4-26.5)	0.09- 1.2 (0.08-1.1)
50 (52.8)	0.9-1.5	9-17 (9.9-18.7)	0.18- 2.3 (0.16-2.1)
100 (105.5)	1.3-2.2	7-13 (7.7-14.3)	0.36- 4.6 (0.33-4.2)
150 (158.3)	1.8-2.8	6-12 (6.6-13.2)	0.54- 6.9 (0.49-6.3)
250 (263.8)	2.6-3.8	5-11 (5.5-12.1)	0.90-11.5 (0.82-10.4)

*Marshall and Swift Cost Index = 561.8 (October 1978).

**No capital related costs.

Note: Retrofit factor = 1.00 - 1.80, dependent on boiler size and age. The smaller and older the boiler, the closer this factor approaches 1.80. This factor should be used to multiply the above capital costs for new plant when retrofitting a limestone scrubber to an existing plant.

Lime Scrubbers

Definition

Lime flue gas scrubbers are gas-liquid contacting devices that use the chemical reactions between $Ca(OH)_2$ and SO_2 to remove the oxides of sulfur from the combustion gases generated during the operation of fossil-fired furnaces.

Application Areas

Lime scrubbers can be used to clean up flue gases from either coal-fired or oil-fired boilers. Though sometimes scrubbers are also expected to remove particulates, the main function of a lime scrubber is to ensure that only a minor amount of SO_2 is allowed to escape with the flue gas. The level of allowed SO_2 emission varies with the local air pollution control regulations; lime scrubbers are used when high levels of SO_2 removal are required and a liquid by-product stream is not permitted. The total amount of solid wastes is generally less than with limestone scrubbers.

Principle of Operation

A simplified flow diagram of a typical lime scrubbing system is shown in Figure 6. Flue gas from the boiler, after passing through a suitable collector for particulate removal, enters the lower part of the scrubber. The flue gas then passes up to a "reaction zone" (which could be a venturi, a system of trays and packing, a spray zone, or a combination of these) where the gas is mixed vigorously with the lime slurry showering down. The gas attains its adiabatic saturation temperature, while the SO_2 is removed according to the overall chemical reaction:

$$Ca(OH)_2 + SO_2 \rightarrow CaSO_3 \cdot 1/2 \ H_2O + 1/2 \ H_2O$$

Some calcium sulfate is also formed, and with forced oxidation, gypsum becomes the primary product.

After this, the flue gas passes through a mist eliminator zone, where the entrained liquids and solids are captured, collected, and washed back down into the scrubbing section. The essentially droplet-free, scrubbed gas may then be reheated, for both buoyancy and stack lining corrosive protection, before venting through a stack into the atmosphere.

The spent slurry, which is a mixture of unused lime, fly ash, calcium sulfite, calcium sulfate, and sulfate solution, is pumped out at the scrubber bottom and back to the reaction tank, where it is mixed with the freshly slaked lime and make-up water. The reaction tank is designed to provide for completion of chemical reactions and precipitation of solids. A bleed stream from the scrubber bottoms is pumped to a settling device (e.g., a clarifier or thickener) and then to the sludge disposal system. The clarified effluent from the settling device is recycled to the lime slurry circuit. A vacuum
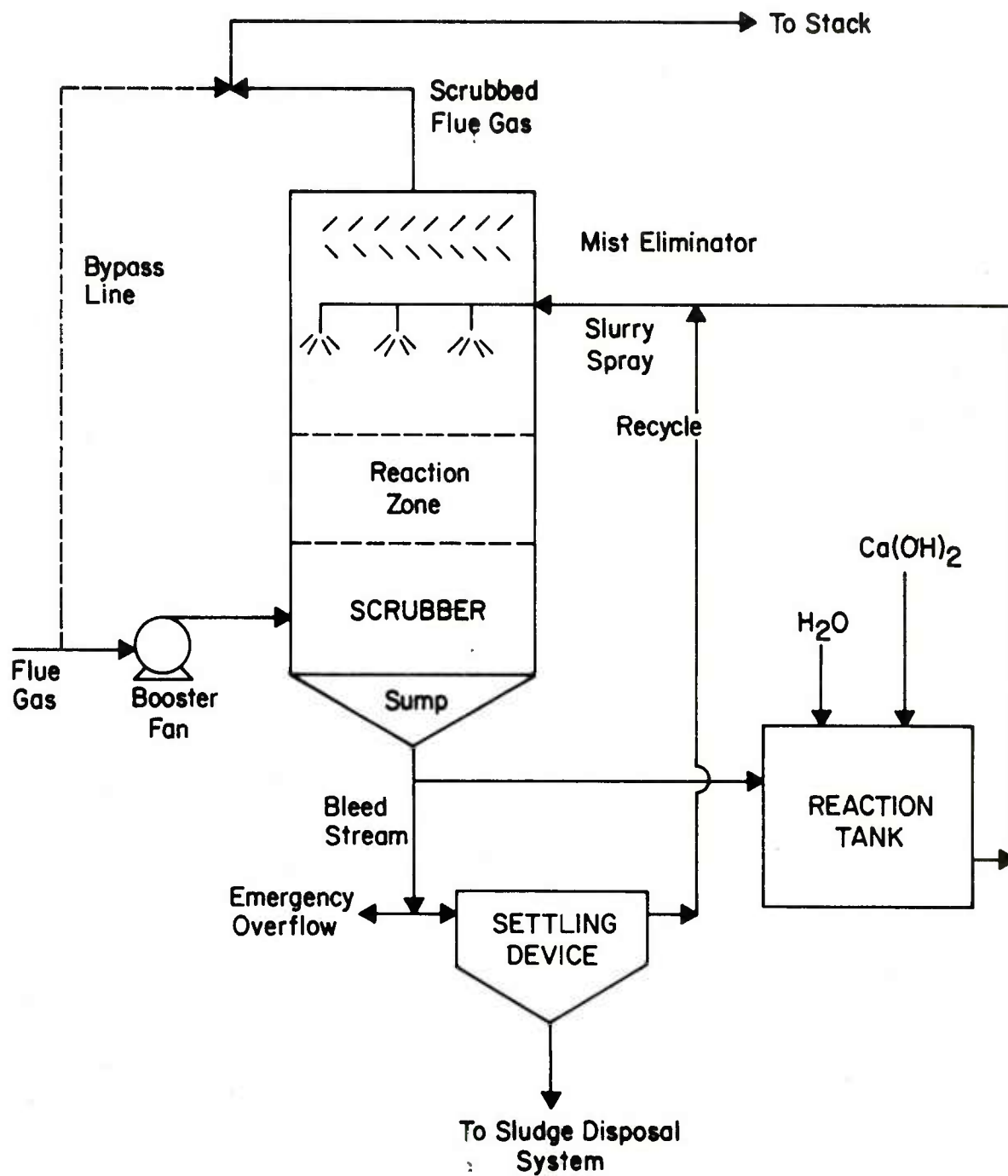
Figure 6. Lime scrubbing system.

filtration system is also used sometimes to treat the waste stream from lime scrubbers.

Advantages and Disadvantages

Lime scrubbers are advantageous because they are relatively simple to operate. Moreover, lime is more reactive than limestone, though it is also more expensive. The cost differential, however, also depends on the shipping distance since lime weighs less than limestone on a molar basis. In addition, the SO_2 removal efficiency of lime scrubbers is generally considerably higher than that of limestone systems on an equivalent cost basis. The lime scrubbing process can tolerate significant fly ash in the system.

Disadvantages of the system include the production of large quantities of solid waste for disposal; the formation of sulfate scale in the scrubber and the demister; corrosion and erosion attack on the pumps, scrubber internals, and reheater tubes; and the safety problems of handling lime.

Efficiency

The SO_2 removal efficiency of lime scrubbers is in the range of about 75 to 99 percent, although the higher end of the range is not normally achieved. Major factors affecting this efficiency include the coal's sulfur content, the stoichiometric ratio of lime to SO_2 used, the liquid-to-gas ratio, and the number of scrubbing stages. The use of lime is also generally very high. As shown in Figure 7, the SO_2 removal is essentially a linear function of the amount of lime employed. Essentially complete removal can be obtained with a stoichiometric ratio of one (Ca/S ratio equal to 1.0) over a range of liquid-to-gas ratios in a two-stage system.

Reliability

In general, the reliability of lime scrubbers is still a problem. Scale formation on ducts and pipes has caused plugging and scaling of the mist eliminators, which leads to excessive mist carryover and the consequent rainout of acid mist from the stack. Closed-loop operation has been difficult because of the large amount of freshwater needed to slake the lime and to prevent plugging of the mist eliminators. The system availability ranges from 40 to 90 percent annually, and can approach 100 percent if carbide lime sludge, a special reagent in limited supply, is used as the scrubbing reagent. Reliability can also be considerably improved if some open-loop operation can be designed into the system -- for instance, by ponding excess liquid as well as solids.

Operational History

No accurate figures on functional life have been recorded for lime scrubbers. Most operating units are still in the demonstration period. The first units in the United States went on stream in 1973. Generally, the
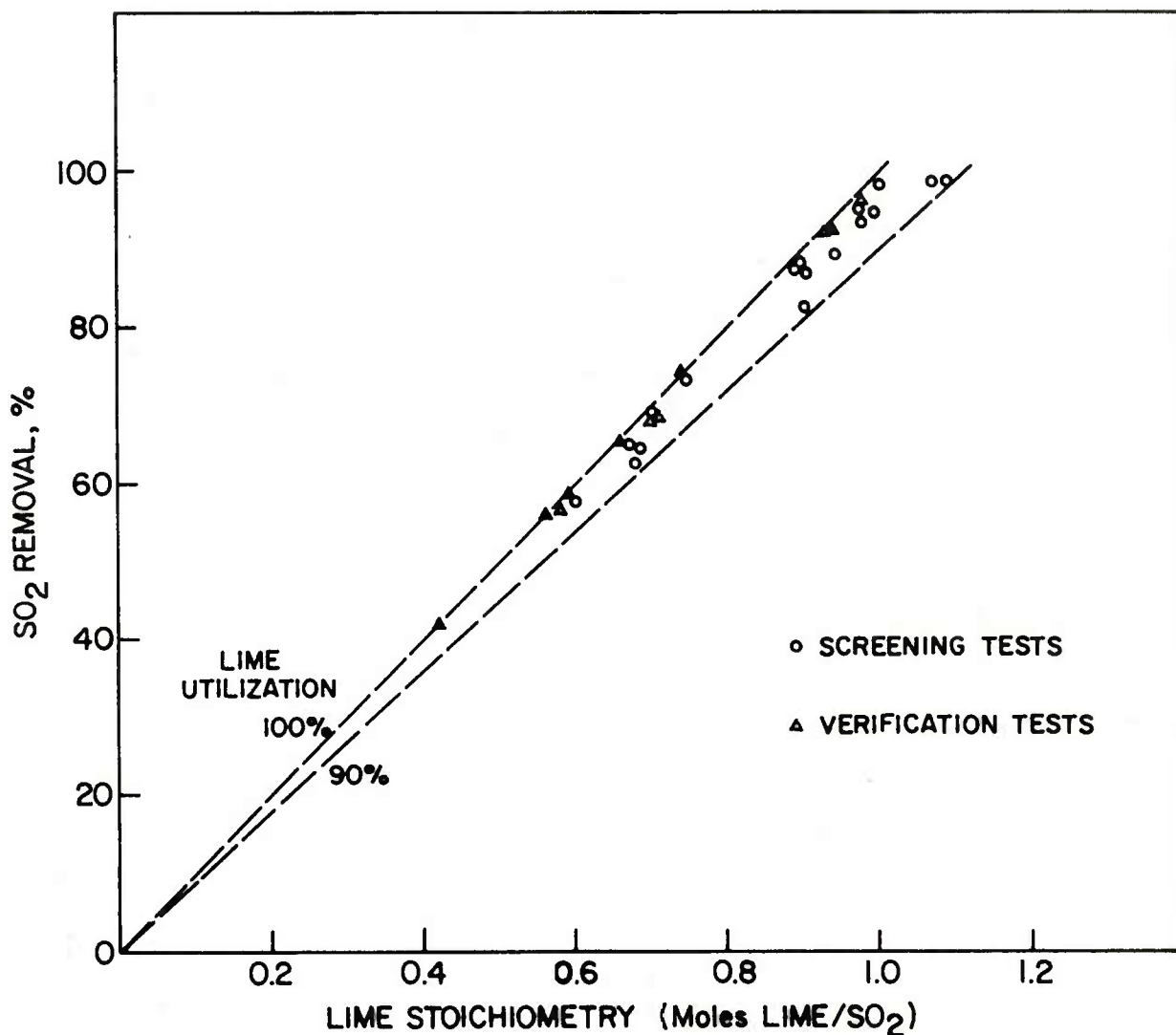
Figure 7. SO₂ removal efficiency as a function of lime stoichiometry in a two-stage contactor. (From E. L. Biedell, R. J. Ferb, G. W. Malamaud, C. D. Ruff, and N. J. Stevens, _EPA Evaluation of Bacho Industrial Boiler Scrubber System at Rickenbacker AFB_, EPA Report 600/7-78-115 [EPA, June 1978].)

higher the level of sulfur in the coal, the more difficult the operation has been to control.

Energy Impact

Lime scrubbers require little energy to operate (averaging about 3 to 6 percent of the plant energy input). The pressure drop of the flue gas through the system will generally not exceed 20 in. (4977 Pa) of water, which is not excessive, but generally does require some added blower or fan capacity at the heating plant.

Factors Affecting Performance

Plugging and scaling are the major problems in using lime scrubbers. With carbide lime as the reactant, the system is essentially scale-free due to the low concentration of sulfate formed; however, carbide sludge is generally unavailable in large quantities. The addition of magnesium hydroxide to the system with commercial lime reduces sulfite oxidation (thus preventing scaling) and dramatically increases the SO_2 removal efficiency. Before widespread use by the Army can be recommended, much more needs to be known both about the chemistry involved in the use of magnesium additives and about the possible environmental impact of these additives.

Retrofit Considerations

Lime scrubbing processes are relatively easy to retrofit because they require much less auxiliary equipment than do regenerative processes. Since bypass capability is often desirable, the scrubbing system can be installed parallel to the ductwork anywhere between the mechanical collector and the stack. In some systems where the scrubber has its own stack, flue gas from more than one boiler can be combined into a manifold duct leading to the scrubbing system separate from the plant. Coarse particulate removal and added fan capacity must normally be included in retrofit plans. Space for waste disposal is generally a major consideration for lime scrubbing systems.

Cost Profile

Table 8 presents the estimated range of capital and recurring cost elements for lime scrubbing systems. Capital costs are installation expenses in FY78 for a lime scrubbing system consisting of a marble bed scrubber, a clarifier, an associated lime-handling system, and a slurry circuit. The costs, however, exclude the sludge pond and flue gas reheat system. The capital cost retrofit multiplier is at the bottom of the table. The operating costs are based on a capacity factor of 80 percent (7008 hours/year) and include recommended operating labor, maintenance, electricity, process water, lime, and waste disposal requirements. Capital-related charges and steam reheat costs are not included in this tabulation.

Table 8

Estimated Costs of New Lime Scrubbing Systems
(Coal of 0.4-4.3 Percent Sulfur)

Boiler Input Heat Rate, 10⁶ Btu/hr (GJ/hr)	Capital Cost ($ Million)*	Operational Cost, $/ton Coal Burned ($/MT)**	Lime, K ton/yr (kt/yr)
25 (26.4)	0.6-1.1	15-24 (16.5-26.5)	0.04-0.5 (0.04-0.45)
50 (52.8)	0.9-1.5	10-17 (11.0-18.7)	0.09-1.1 (0.08-1.00)
100 (105.5)	1.3-2.2	8-12 (8.8-13.2)	0.17-2.2 (0.15-2.00)
150 (158.3)	1.8-2.8	7-11 (7.7-12.1)	0.25-3.2 (0.23-2.90)
250 (263.8)	2.6-3.8	6-10 (6.6-11.0)	0.42-5.4 (0.38-4.90)

*Marshall and Swift Cost Index = 561.8 (October 1978).

**No capital related costs.

Note: Retrofit factor = 1.00 - 1.80, dependent on boiler size and age. The smaller and older the boiler, the closer this factor approaches 1.80. This factor should be used to multiply the above capital costs for new plant when retrofitting a lime scrubber to an existing plant.

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Double-Alkali Scrubbers

General Description

Double-alkali flue gas scrubbers involve the same chemical reactions between soluble alkali -- such as NaOH or Na_2CO_3 -- and SO_2 to remove the oxides of sulfur from the combustion gases as do single-alkali processes. However, the resulting scrubbing liquor from the double-alkali scrubber is then directed to a regeneration system, where sodium is regenerated by reaction of the spent liquor with lime, producing insoluble calcium-sulfur waste salts, hence the name double alkali. At least in theory, no liquid stream (other than the liquor adhering to the washed solids) needs to be purged from the system.

Double-alkali scrubbers can be used to clean up flue gases from either coal-fired or oil-fired boilers. Generally, it is desirable to use this process on relatively particulate-free flue gas. The main function of the scrubber is to ensure that the amount of SO_2 in the treated flue gas meets local air pollution control regulations when generally high removal levels are required. Double-alkali systems may be operated in either a dilute or concentrated mode. Applications of the concentrated mode process include high-sulfur coal burning in pulverized-coal-fired boilers, where there is sustained operation with low excess air. The ratio of oxygen to SO_2 is important to this mode in order to limit the amount of oxidation of sulfite to sulfate, and to prevent scaling in the scrubber. The oxidation of absorbed SO_2 should be less than about 25 percent.

The dilute mode process is usually applicable when the flue gas has a high concentration of oxygen relative to the SO_2 concentration (e.g., with low-sulfur coal, high excess air firing). The dilute mode process can also better tolerate fluctuations of flue gas composition due to boiler upsets or a change to a different composition fuel.

Technical Description

A simplified diagram of a typical dilute mode double-alkali scrubbing system is shown in Figure 8. The schematic flow is practically the same for a concentrated mode process, except that the clarifier and the level control tank are not used, and the soda ash mix tank also serves as a recycle tank. Freshwater make-up, thickener overflow, and soda ash are mixed together in this tank and the resulting liquor is returned to the scrubber top. The concentrated mode process thus tends to be more compact and requires less space than does the dilute mode process.

In general, the double-alkali process consists of three major steps: absorption, chemical regeneration, and solids dewatering. During absorption, dirty flue gas from the boiler, after passing through a suitable collector for particulate removal, enters the lower part of the scrubber. The flue gas then passes up to a "reacting zone" (which could be a venturi, a system of trays

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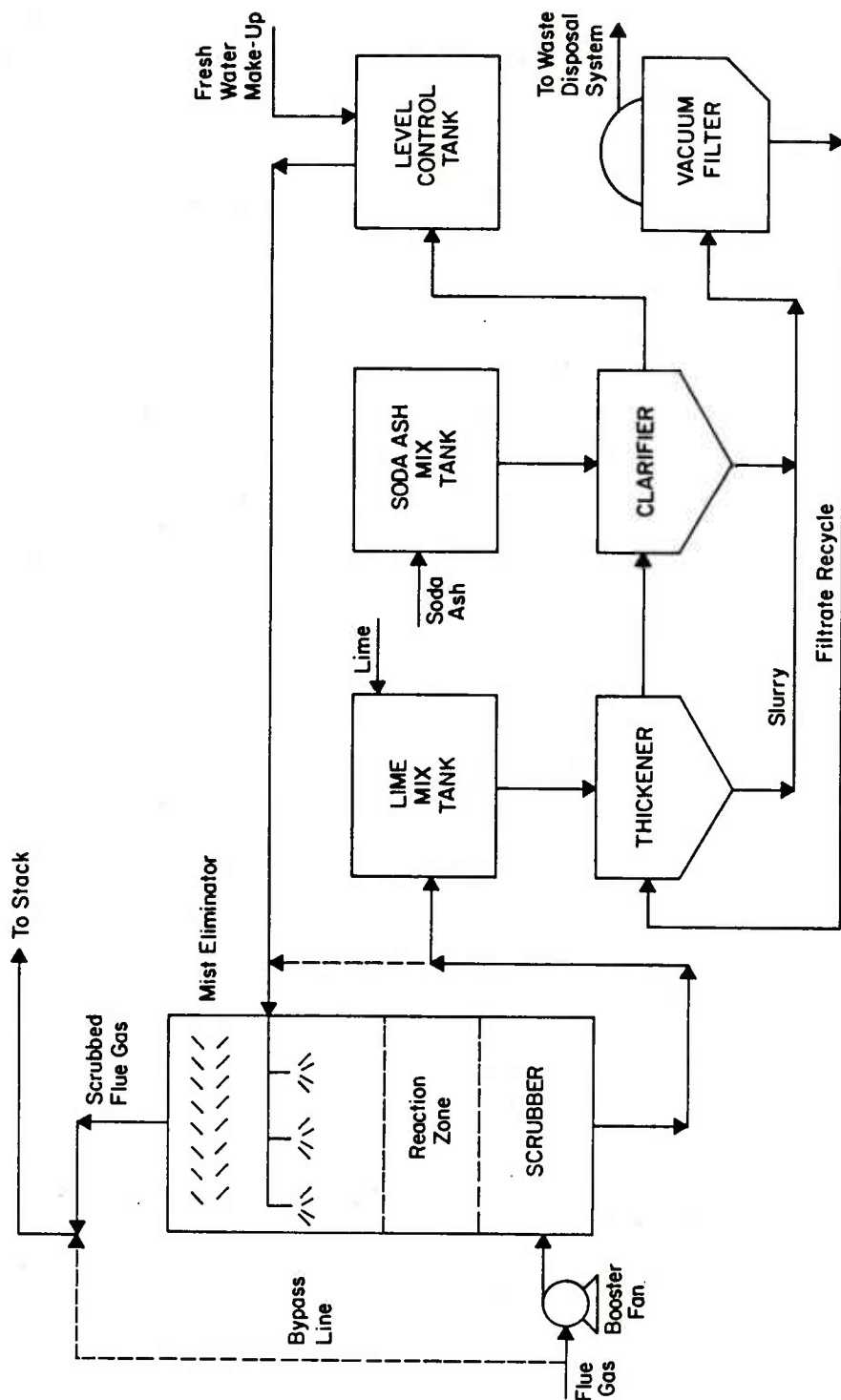
Figure 8. Dilute mode double-alkali scrubbing system.

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and packings, a spray zone, or a series or combination of these), where the gas is mixed vigorously with the alkali solution of sodium hydroxide and sodium sulfite (fed from the recycle or level control tank), showering down. The gas attains its adiabatic saturation temperature, while the SO_2 is removed according to the following reactions:
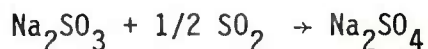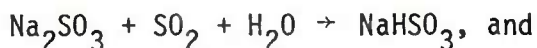
$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2,$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow NaHSO_3, \text{ and}$$

$$Na_2SO_3 + 1/2\ SO_2 \rightarrow Na_2SO_4$$

After this, the flue gas passes through a mist eliminator zone, where the entrained liquids (and dissolved solids, if any) are captured, collected, and washed back down into the scrubbing section. The essentially droplet-free, scrubbed gas may then be reheated, for both buoyancy and stack lining corrosive protection, before venting through a stack into the atmosphere.

During regeneration, the spent scrubbing solution from the scrubber bottom is converted to disposable sulfur-bearing solids and reusable sodium-based scrubbing liquor. The primary raw material used for this is lime, which reacts with Na_2SO_3, $NaHSO_3$, and Na_2SO_4 in the scrubber effluent according to the reactions:

$$Na_2SO_3 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_3$$

$$2NaHSO_3 + Ca(OH)_2 \rightarrow Na_2SO_3 + 2H_2O + CaSO_3$$

$$Na_2SO_4 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_4$$
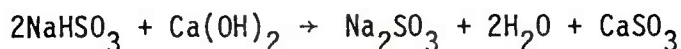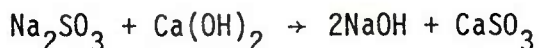
Na_2CO_3 is also added to replenish sodium losses and to further prevent scaling.

In solids dewatering, the thickened slurry resulting from chemical regeneration is dewatered in a vacuum filter and forms an inert cake which can be easily disposed of. The cake is also washed to minimize sodium losses, and the filtered liquor is recycled to the system.

The concentrated mode double-alkali scrubbing system has a minimum level of fuel sulfur content, below which it cannot be operated without an intentional purge of Na_2SO_4. This lower limit is a function of the rate of oxidation of sulfite to sulfate in the system and is therefore also dependent on the level of excess air applied in firing. In the dilute mode process, a large excess of sulfate ion in the regeneration section is essential to establish a chemical equilibrium favoring caustic formation. The high oxidation rate, however, may contribute to poor lime use. Carbonate softening with soda ash is necessary to prevent scrubber plugging. Soda ash, however, affects

sodium losses in the filter cake and requires careful control if the system is to be operated in a closed-loop mode.

Process Characteristics

The first double-alkali units in the United States went on stream in 1974. Most, if not all, are still in the demonstration phase of operation. Because the dilute mode process is not constrained by an upper level of SO_2 oxidation, it is more widely used than the concentrated mode; in fact, high oxidation rates usually improve the properties of the waste sludge in the dilute mode. The dilute phase process may also be easier to operate, but this has not been adequately demonstrated to date.

Double-alkali scrubbers require little energy to operate, averaging about 2 to 3 percent of the plant energy input. The pressure drop of the flue gas through the system will generally not exceed 10 to 15 in. (2488 to 3733 Pa), which is not excessive, but obviously requires added energy for fans. Vacuum pumps are also needed on the filter. Because two liquid circuits are involved, the system requires extra pumps; for this reason, the double-alkali process is probably slightly more energy intensive than the single-loop processes.

The SO_2 removal efficiency of double-alkali scrubbing systems is generally high, in the range of about 85 to 95 percent. Major factors affecting this efficiency include the coal's sulfur content, the lime and sodium stoichiometry used, number of scrubbing stages, and type of scrubber, as well as the operating mode and the liquid-to-gas ratio. The removal efficiency is in the same range as the single-alkali process because the chemistry in the first part of the double-alkali process is the same as that involved in the single-alkali system. (Data obtained in a double-alkali scrubber that employed a venturi and then two-tray stages are shown in Figure 9.) It is more difficult to remove high percentages of SO_2 from low SO_2 content flue gas than it is to remove the same percentage from flue gas containing higher levels of SO_2. Obviously, higher levels of available sodium in the scrubber liquor will permit higher levels of SO_2 removal. The overall process, of course, must be designed so that the regeneration reactions in the liming equipment proceed at a rate which equals or exceeds the rate at which SO_2 in the flue gas reacts with the soluble alkali in the scrubber.

Because the double-alkali scrubber itself is relatively free from plugging and scaling problems, its reliability is high. However, in most units currently operated, the system availability has been rated at an average from about 70 to 80 percent because components other than the scrubber can accumulate scale, and frequently do. During favorable operating conditions, a level near 100 percent availability has been approached, and perhaps such higher levels can be anticipated in the future if personnel with technical expertise operate the process.
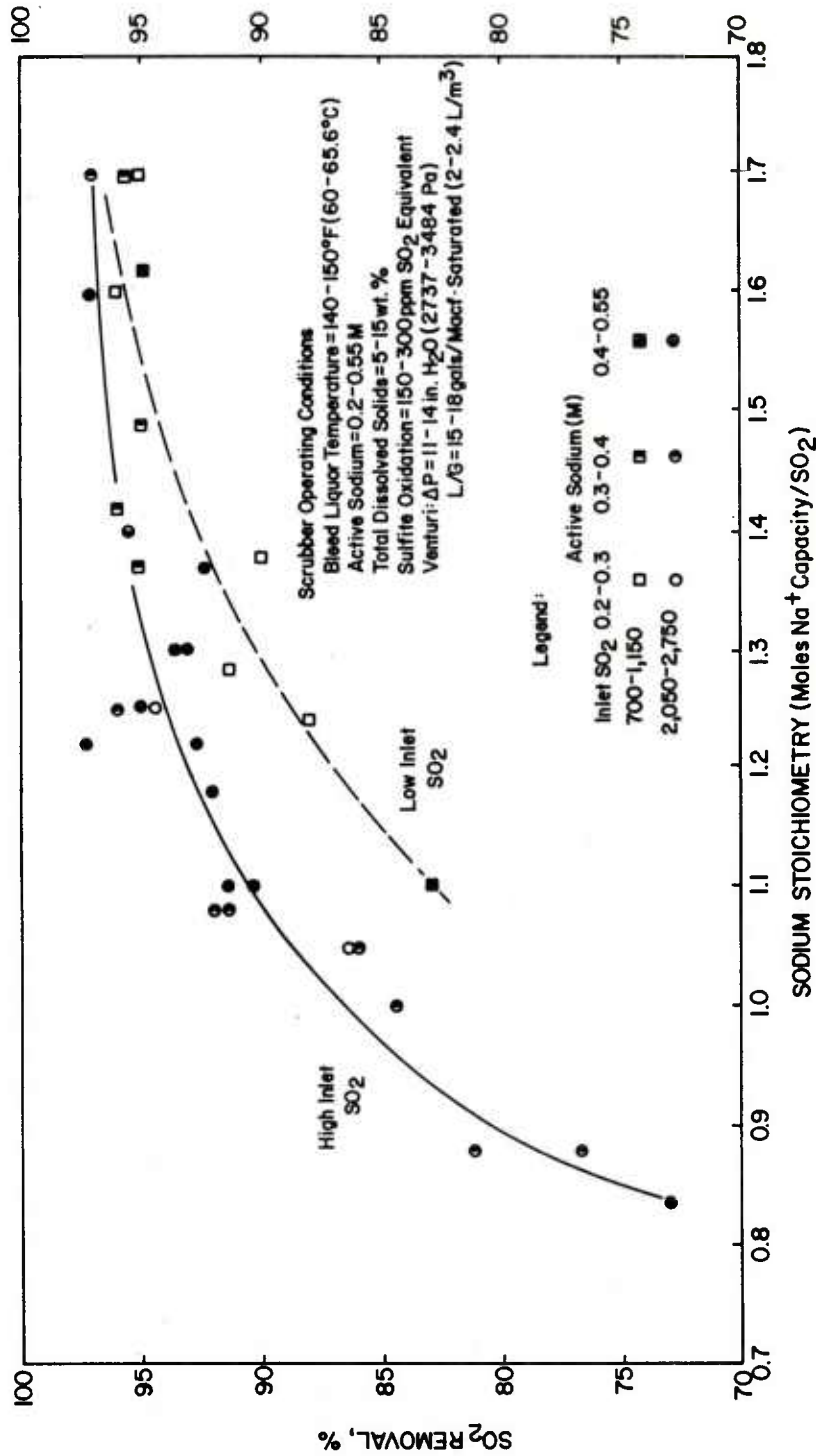
Figure 9. SO_2 removal efficiency as a function of feed stoichiometry in a venturi and two-tray, double-alkali contactor. (From C. R. LaMantia, R. R. Lunt, J. E. Oberholtzer, E. L. Field, and J. R. Valentine, Final Report: Dual Alkali Test and Evaluation Program Volume II. Laboratory and Pilot Plant Programs, EPA Report 600/7-77-050b [EPA, May 1977].)

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The major advantage of double-alkali scrubbing systems is that they present few scrubber plugging and scaling problems, since no solids are formed by the SO_2 absorbing reactions. The highly reactive reagent used by the process also permits a low operating liquid-to-gas ratio in the scrubber. A relatively pure gypsum by-product is produced and can be landfilled at most installations. It is expected that fewer corrosion and erosion problems will be encountered, but this feature has probably not been satisfactorily demonstrated to date.

The process, however, has some disadvantages as well. It is relatively complex, compared with other SO_2 scrubbing systems such as lime, limestone, or single alkali. The cost of reagents used may be somewhat higher. Scale formation in the regeneration loop has caused operating problems, and a good control system is essential to reliable operation. The filter cake produced may contain significant amounts of sodium salts (particularly under upset conditions), which may cause environmental concern in some locations. At several industrial sites, there have been high particulate emissions from the scrubber due to misting. Adequate protection for the health and safety of workers handling both caustic and lime is another important design and operation consideration.

The double-alkali process may be somewhat more difficult to retrofit than the simpler FGD systems because of the extra auxiliary equipment involved. However, because bypass capability is often desirable, and since FGD is usually one of the final processes in a power plant, most FGD equipment can be housed in a large multilevel structure or parallel to the bypass ductwork. Because of the relatively large number of vessels used to process the various aqueous solutions and slurries, special precautions against freeze-ups must be considered in areas where low temperatures are common. Coarse particulate removal and added fan capacity must normally be included in retrofit plans. And solid wastes must be moved to a landfill site, so availability of transportation must also be considered.

Magnesia Scrubbers

General Description

Magnesia flue gas scrubbers are gas-liquid contacting devices that use the chemical reactions between MgO and SO_2 to remove the oxides of sulfur from the combustion gases. The higher the level of sulfur, the more attractive the process, because SO_2 is a marketable by-product. The scrubber produces magnesium sulfite and sulfate, which are then further processed to make relatively concentrated SO (e.g., for H_2SO_4 production) and MgO for reuse. Because the reactant, MgO, is recovered by this second step, the process is

classified as a regenerative process, and no significant solid waste is produced.

Magnesia scrubbers can be used to clean flue gases from either coal-fired or oil-fired boilers to relatively high efficiency levels. For coal-fired applications, highly efficient particulate removal is required to avoid large-scale contamination of the magnesia slurry with fly ash. Because the slurry solids are continuously recycled to the scrubber after each regeneration, even small amounts of ash continuously accumulating in the slurry will eventually choke the system.

Although not tested in commercial use yet, this SO_2 removal technology should be valuable in geographical areas near sulfuric acid markets, and where there is a limited amount of land for the disposal of sludges produced from throwaway FGD processes. The process should also be of special interest to chemical or fertilizer companies which have their own power generation system, and to users of sulfuric acid. The process is not of interest to potential buyers with small isolated boilers because the regeneration plant is a major chemical operation in itself.

Technical Description

A simplified flow diagram of a typical magnesia scrubbing system is shown in Figure 10. Flue gas from the upgraded fly ash collectors enters a prescrubber where it is sprayed with water for residual fly ash removal and for adiabatic cooling to about $125^\circ F$ ($51.7^\circ C$). Various highly efficient particulate removal devices could probably be employed. Next the gas enters the SO_2 absorber (in this case, a venturi scrubber), where it is met with a circulating aqueous slurry containing MgO, magnesium sulfite ($MgSO_3$), and $MgSO_4$. Sulfur dioxide is removed from the gas according to the following reactions:

$$MgO + SO_2 + 3H_2O \rightarrow MgSO_3 \cdot 3H_2,$$

$$MgO + SO_2 + 6H_2O \rightarrow MgSO_3 \cdot 6H_2O, \text{ and}$$

$$MgSO_3 + 1/2\ O_2 + 7H_2O \rightarrow MgSO_4 \cdot 7H_2O$$
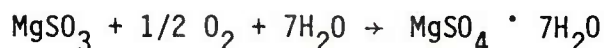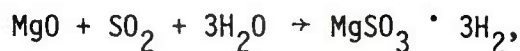
The scrubbed gas then passes through a system of mist eliminators for removal of entrained materials, and the cleaned gas may be reheated before discharging to the stack.

The scrubber effluent is directed to the mother liquor tank for recycle. A side-stream is continuously withdrawn to the solids separation system so that the solids concentration in the absorber bottoms is held at about 10 percent. In the solids separation system -- which may be a thickener, a centrifuge, or both -- the scrubbing slurry is reduced to a cake containing crystals of hydrated magnesium-sulfur salts and unreacted MgO. The recovered liquor can then be returned to the mother liquor tank or sent to an MgO recovery
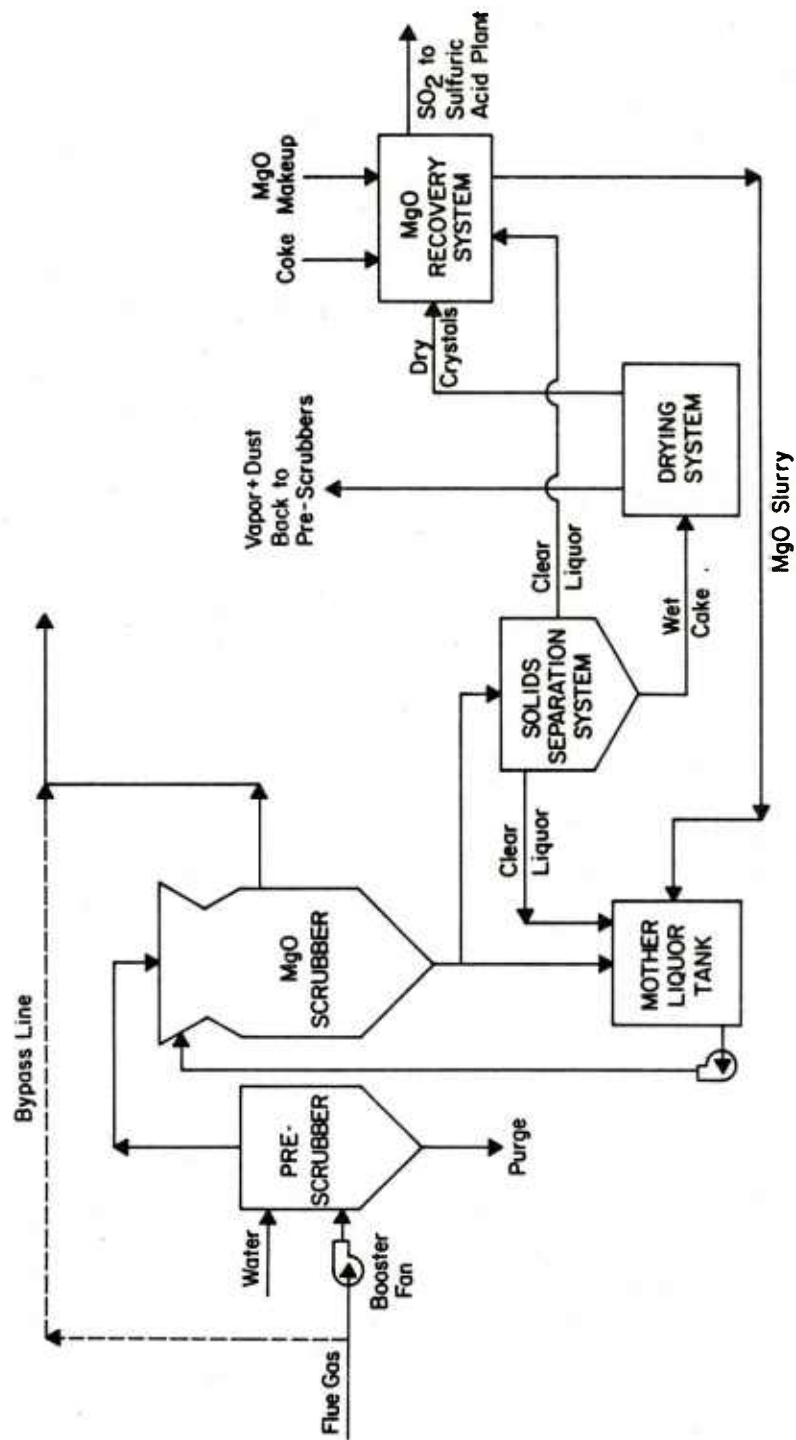
Figure 10. MgO scrubbing system.

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system, where it will help in quenching the regenerated MgO coming from the calciner. The cake from the solid separation system is then dried -- thus removing free and bound moisture, which is recycled to the prescrubber. The anhydrous magnesia crystals from the dryer are next sent to the MgO recovery system, where coke and MgO make-up are added, and MgO and SO_2 are regenerated according to the reactions:

$$MgSO_3 \xrightarrow{Heat} MgO + SO_2, \text{ and}$$

$$MgSO_4 + C + 1/2 \ O_2 \rightarrow MgO + SO_2 + CO_2.$$
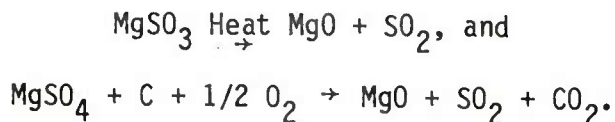
Sulfur dioxide leaving the recovery system is finally sent to the H_2SO_4 manufacturing plant, and the MgO is quenched and slurried before recycling to the mother liquor tank.

Process Characteristics

No accurate figures on functional life have been recorded for magnesia scrubbers. These systems have operated on utility boilers since about 1972. But only one is still working (Potomac Electric Power's Dickerson Station), and this unit has run about 50 hours during the past two years. The other units have now been permanently shut down due to various difficulties not only with the plants themselves, but also with the commercial arrangements for regeneration of the magnesium sulfite-sulfate intermediate product.

Although drying and calcining operations are performed in the magnesia scrubbing process, the energy requirements of this system are still relatively modest (averaging about 4 to 7 percent of the plant energy input). The pressure drop of the flue gas through the system will generally not exceed 20 in. (4977 Pa), which is not excessive. With all wet scrubbing processes, energy requirements for reheat are relatively process independent.

The SO_2 removal efficiency of magnesia scrubbers is in the range of 92 percent or higher. Major factors affecting the scrubbing performance include the inlet SO_2 concentratration (or fuel sulfur content), scrubber pressure drop (which is somewhat dependent on the liquid-to-gas ratio), the recycle slurry pH, and its solid contents. The effect of slurry pH is shown in Figure 11. The rate of MgO addition controls the slurry pH, and the slurry bleeding rate adjusts the slurry solid contents. Generally, as the pressure drops more, SO_2 removal efficiency increases, and inlet SO_2 concentration influences removal efficiency less.

Although no fundamental problems involving scrubber chemistry have been observed to date, the reliability of MgO scrubbers has been erratic due to their structural complexity. In fact, MgO scrubbers, on the average, seldom work; average availability may be rated at close to zero, although occasionally a system operates more than 80 percent of the time in a given month. The most serious operating problems have been caused by the solids-handling system -- e.g., excessive dusting, solids buildup, inefficient drying. Although the
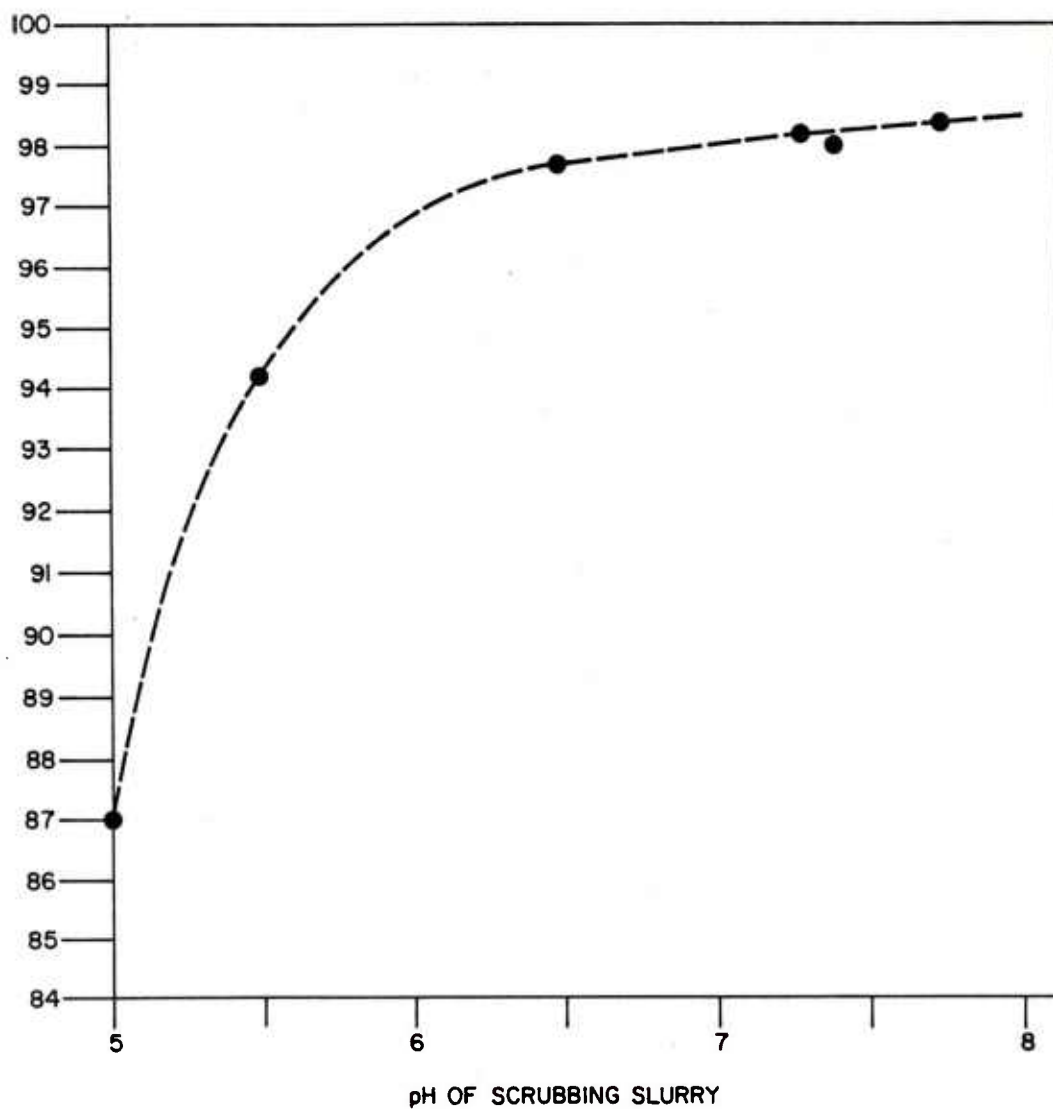
Figure 11. The effect of pH on SO_2 removal efficiency in MgO scrubbing. (From B. M. Anz, G. C. Thompson, Jr., and J. T. Pinkston, "Design and Installation of a Prototype Magnesia Scrubbing Installation," paper presented at the EPA Symposium, New Orleans, LA, May 1973.)

scrubber itself is free from scaling and plugging, the MgO slaker is easily plugged when recycle mother liquor is used for slaking. The use of freshwater in the slaker, however, may cause excessive MgO loss to the blowdown stream. Particle size of regenerated MgO is also an important factor to the slaking operation; for instance, a pulverizer may be needed. Rubber or plastic lining is considered necessary in slurry-handling applications (pumps, valves, piping) to prevent erosion and corrosion. Air leakage in the calciner may also cause operating problems. The calciner must operate near neutral or reducing conditions in order to efficiently reduce magnesium sulfate to magnesia.

Process Summary

One of the major advantages of the magnesia scrubbing process is that it produces no major waste streams other than fly ash. It is also free from scaling and plugging problems; and due to the very active reagent used, a high SO_2 removal efficiency is possible. Oxidation is tolerated in the system, and with its capability of producing high quality sulfuric acid, part of the operating cost can be recovered by sale of H_2SO_4.

Disadvantages, however, are many. First, the system cannot tolerate fly ash (at least the ash must be kept out of the MgO regeneration system), and requires a relatively high liquid-to-gas ratio in the scrubber. The system involves complex solids handling operations, which sometimes cause excessive dusting, buildup of solids, and loss of capacity in the drying section. Also, auxiliary fuel is required for the drying and calcining operations. Losses and slight deactivation of MgO may result after repeated regeneration, and the cost of this reagent is high. Slurry carryover by the flue gas may also be troublesome. In addition, the availability and dependability of an existing sulfuric acid plant and market is essential for this FGD process to become commercially practical.

The ease or difficulty of retrofitting a magnesia scrubbing system is obviously highly site-specific. However, even if an existing sulfuric acid plant is located nearby, substantial space may be needed to accommodate the equipment involved in the regeneration system. Retrofitted magnesia scrubbing systems do not seem to be an attractive option -- except in geographical areas where waste disposal is a problem, or where there is a ready and dependable use of H_2SO_4. The size of the installed heating or steam plant (or group of such plants) must also be relatively large to make this scrubbing system economically justifiable.

Citrate Scrubbers

General Description

Citrate flue gas scrubbers capitalize on the buffering effect of a sodium citrate solution to improve the aqueous absorption of SO_2 from the gases

generated during the operation of boilers and other industrial processes. The absorbed SO_2 subsequently reacts with H_2S gas to precipitate elemental sulfur and release citrate ions for reuse. The citrate process is therefore a regenerable process.

The citrate process is adaptable to virtually all sources of sulfur-bearing industrial gases due to its insensitivity to rapid fluctuations in load or SO_2 concentration. It can efficiently treat flue gases with SO_2 concentrations ranging from over 50,000 ppm to less than 2000 ppm (by volume), and therefore is being considered for use with boilers and other industrial operations, such as smelters. Flue gas quenching and particulate removal are required before citrate scrubbing.

The economics of this scrubbing process become most attractive in two cases: (1) when the amount of SO_2 to be removed increases (perhaps because of high-sulfur fuel firing, strict emission regulation, or the large size of the power plant); and (2) when the option to recover sulfur becomes desirable (because land for waste disposal is not available, or sulfur is in demand). The system has a minimal environmental impact because there is no significant sludge disposal requirement. An acid bleed stream from the prescrubber, and a small amount of Glauber's salt ($Na_2SO_4 \cdot 10 H_2O$) produced in the process must be neutralized and either disposed of or sold. The citric acid reagent itself is biodegradable organic material contained in foods and poses no serious environmental problem.

Technical Description

A simplified flow diagram of the citrate scrubbing system is shown in Figure 12. Flue gas, before entering the scrubber, passes through an ejector-type venturi prescrubber, or similar device, for quenching and removing of residual dust particles, SO_3, and HCl mists. Transfer of SO_2 to the citrate solution in the scrubber should be more efficient if the flue gas is first subcooled below its adiabatic saturation temperature. A bleed stream is taken from the venturi recycle stream to prevent buildup of ash and acids. This stream is neutralized and sent to the ash pond.

The cooled gas then enters a packed absorber (preferred for flue gases derived from coal combustion, although trayed absorbers or similar devices might also be used), where it flows up countercurrent to a descending solution of sodium citrate, citric acid, and sodium thiosulfate. The SO_2 is essentially absorbed as the bisulfite and controlled by the following reaction:

$$SO_2 + H_2O \rightarrow HSO_3 + H^+$$

The treated flue gas leaving the scrubber must usually be reheated before being discharged to the atmosphere. Citrate solution, containing a bisulfite-citrate complex from the scrubber, is pumped to a series of reactors in the sulfur-precipitation section, where the bisulfite complex reacts in
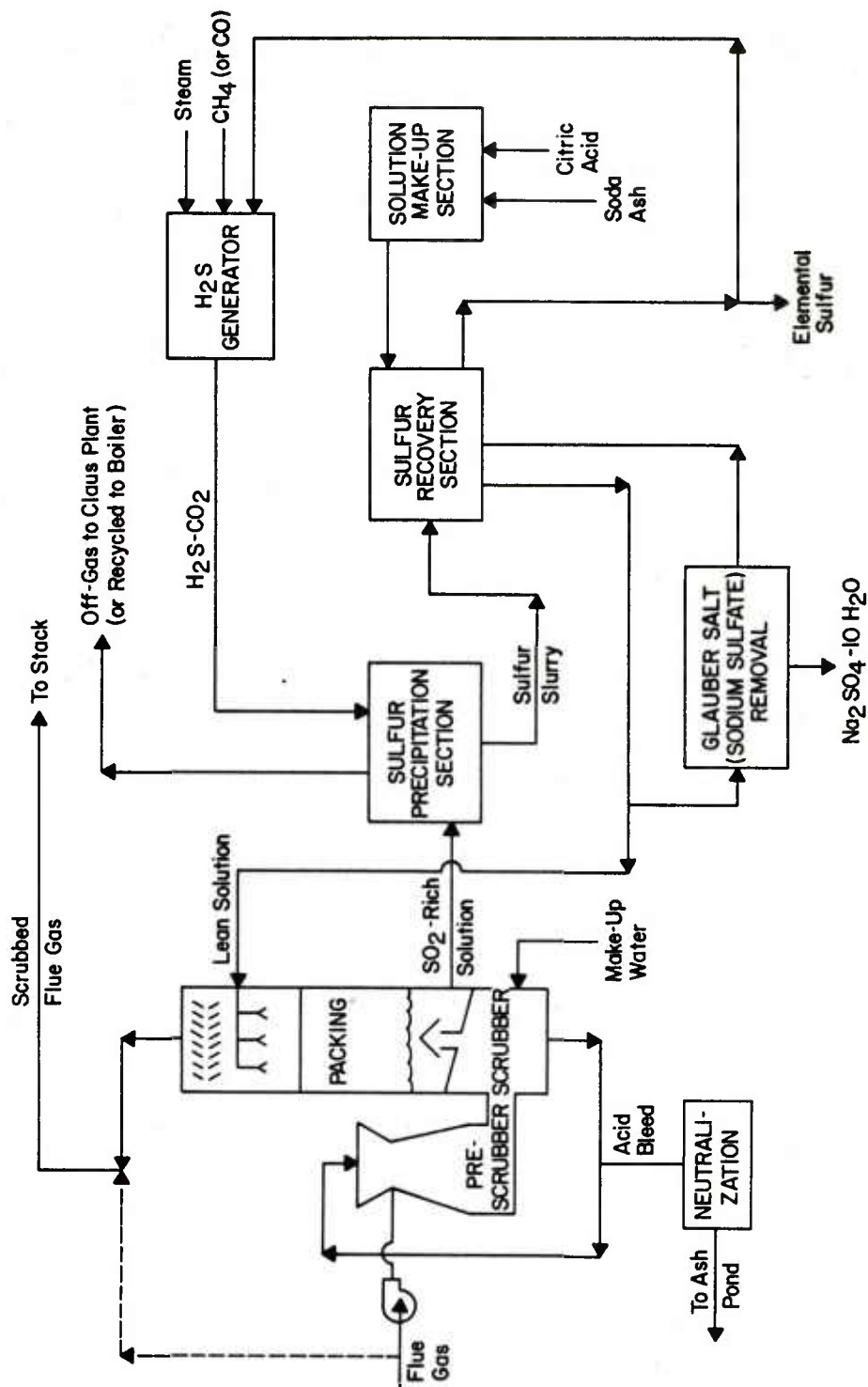
Figure 12. Citrate scrubbing system.

liquid phase with H_2S to form elemental sulfur, thereby regenerating the citrate solution, which is recycled to the scrubber.

The sulfur slurry from the precipitation section is next sent to a sulfur recovery section, where the slurry is heated with steam, and molten sulfur separates in a pressurized, steam-jacketed decanter vessel. Make-up solution may also be added to the circulating citrate solution; these solutions are recycled to the scrubber, after the Glauber's salt is removed in a crystallization unit. The elemental sulfur produced (usually in excess of 99.5 percent purity) may be sent to an acid plant, used on site for H_2S generation, sold, or landfilled.

Process Characteristics

Although the citrate process was conceived and developed by USBM in the late 1960s, the pilot-plant stage -- on a base metal smelter -- was not completed until 1976. A 156,000 scfm (73.6 m^3/sec) retrofit scrubbing facility was scheduled for completion in early 1979. After start-up and acceptance testing, USBM has planned a 1-year demonstration program for comprehensive emission testing and performance evaluation. The energy required by the citrate scrubbing system should be relatively small (averaging about 3 to 5 percent of the plant energy input). The pressure drop of the flue gas through the system will probably not exceed 10 in. (2488 Pa) of water (design figures stated as 4 to 6 in. [995 to 1493 Pa] for the scrubber), which is quite low compared with other, more popular FGD scrubbing systems. Reheat, of course, may result in further relatively small energy penalties.

The SO_2 removal efficiency of citrate scrubbing systems should be very high. More than 90 percent of SO_2 removal can be achieved easily. Efficiencies near 100 percent have been documented in the pilot-plant stage. The factors affecting commercial performance are not yet fully defined, but the effects of temperature and citrate concentration at relatively high SO_2 levels are shown in Figure 13. From the pilot-plant experience, it appears that sub-cooling of the flue gas is necessary to ensure a high level of SO_2 removal. The lean citrate solution recycled to the scrubber should also be slightly warmer than the flue gas so that water will evaporate from the liquor, thus maintaining a water balance for the overall plant. Besides temperature, pH is another major factor influencing the SO_2 absorption. It may be necessary to minimize the level of unreacted H_2S in the sulfur precipitation section (Figure 12). The presence of unreacted H_2S lean citrate solution will form a colloidal sulfur and appear cloudy. Presumably, the sulfur content of the flue gas and the liquid-to-gas ratio are also major operating parameters.

No information on the process' reliability has been reported because construction of the first commercial scrubbing system was scheduled to be completed in mid-1979.
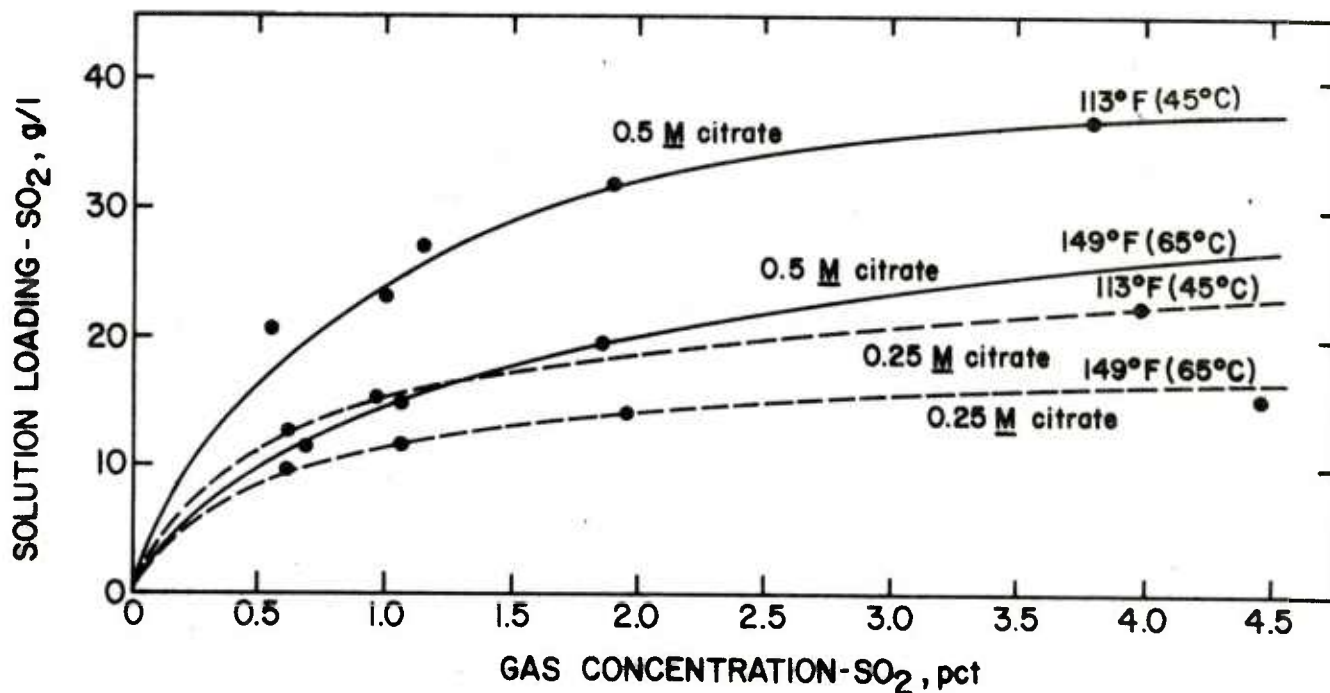
Figure 13. Effect of gas concentration, solution temperature, and citrate concentration on SO_2 absorption. (From J. B. Rosenbaum, W. A. McKinney, H. R. Beard, Laird Crocker, and W. I. Nissen, Sulfur Dioxide Emission Control by Hydrogen Sulfide Reaction in Aqueous Solution, RI 7774 [U.S. Bureau of Mines, 1973].)

Process Summary

The citrate process is supposed to offer many advantages, although in reality these claims have not been demonstrated in commercial operation. The process has no scaling and plugging problems, has a high capacity for short-term SO_2 overloads, and can accommodate, with high efficiency, rapid load changes and SO_2 concentrations; it has a low operating liquid-to-gas ratio (about 7 to 8 gal/1000 scf [0.94 to 1.07 L/m^3]), and a low demand on energy in terms of both process steam and electrical power. Furthermore, marketing the by-product, elemental sulfur, should be less costly than the storage and transportation requirements associated with H_2SO_4. The high purity sulfur product may be of feedstock quality for many industrial processes. There should also be no major waste disposal problem, and environmental impact should therefore be minimal. Disadvantages, however, include the many raw materials required: lime for the acid bleed neutralization; soda ash and citric acid for the scrubbing itself; and highly toxic hydrogen sulfide for sulfur recovery. In addition, a prescrubber is needed to remove all residual fly ash. Whether the process is economical, of course, still needs to be proven.

Many factors must be considered before a citrate scrubbing system is retrofitted to an existing plant. First, unless significant area is available, retrofitting may be difficult due to the structural complexity of the system. However, since the regeneration and scrubbing operations are physically separated, the space requirements can be overcome to some extent by performing regeneration operation somewhere other than at the emission source. The availability of on-site reduction gas feedstocks, as well as the existence of a ready sulfur market must also be considered. Other considerations, as in the case of all FGD processes, include space for bypass duct work, improved particulate control, instrumentation, as well as the size of the scrubber module itself.

Although installation personnel usually do not select FGD system or equipment, their knowledge of each system will help advise the Chief of Engineers in selecting a reputable architect-engineer who will design/select equipment that will satisfy installation design criteria. No single FGD system will be best for every installation. Therefore, the guidelines in this chapter were developed to help installations select equipment meeting their particular needs. Potential purchasers of a system should remember, however, that nontechnical factors (e.g., site specifics, guarantees, waste options) may override the purely technical factors (e.g. economics, efficiency) on which investment decisions are usually based.

Suppliers of FGD equipment often prepare one or more specific designs from information supplied by the installation which will use the system (Figure 14). Thus, the user often must decide: (1) the chemical process to be employed, (2) the gas-liquid contactor to be used in the selected chemical process, (3) the type of treatment the flue gas will receive after it exits the scrubber, and (4) the method of disposal of solid waste products.

Selection of Chemical Process

CERL recommends that military installations initially consider the chemical processing schemes discussed in Chapter 3: (1) single-alkali, (2) lime, (3) limestone, (4) double-alkali, (5) citrate, and (6) magnesia scrubbing. These six processes are listed in order of increasing technical complexity and investment costs, and in decreasing order of reactant costs. The net effect is that over the life of the system, annual costs tend to equalize; thus, other factors will weigh heavily in the decision-making process.

These considerations include extent of process use; the percentage of SO_2 removal required; the space available for the equipment at the power plant; the desired reliability of the equipment; the process' simplicity; the ease of modifying the process after installation (the process' flexibility); the number of waste disposal options available once the chemical process has been selected; and the initial, annual, and life-cycle costs. These factors have been examined for each of the six chemical systems and ranked in terms of average, better-than-average, or worse-than-average performance in Figure 15. If any factor were critically important at a given site, a system ranking at 3 -- or possibly 2 -- on that factor probably would not be suitable.

Extent of Use

The extent of use measures how widely the process has been accepted. Extensive use does not assure that the process represents the best technology, but does indicate that information is available on performance characteristics. Thus, difficulties can be predicted and maintenance planned. The

1. Code requirements
   (Please provide this information in one of the following--lbs/MBTUs, grs/SCFD, or lbs/hr)

	EPA	STATE	LOCAL
Today			
particulate	___	___	___
gaseous	___	___	___
opacity	___	___	___
Foreseeable Future			
particulate	___	___	___
gaseous	___	___	___
opacity	___	___	___

2. Required compliance date:_____

3. Procedure required for compliance testing:_____

   a. Do you conduct your own tests? _____

4. New emission control system, retrofit, or conversion?_____

5. Which type of firing?

   _____ Pulverized coal   _____ Spreader Stoker
   _____ Ash reinjection   _____ Chain Grate Stoker
   _____ Multiple   _____ Other (Specify)

6. Boiler Make & Model _____ Age: _____
   Expected boiler years remaining:_____

7. Maximum continuous rating:_____ lb/hr steam

8. Normal load _____ lb/hr steam
   turndown _____ lb/hr steam

9. Mode of operation: (check one)
   _____ continuous
   _____ intermittent   What cycle? _____
   _____ reserve
   _____ shutdown
   _____ number of operating days per year

10. Design pressure: _____PSIG   Operating pressure _____ PSIG

11. What are the primary fuels and expected percent usage?

    __ __ % bituminous coal   __ __ % bark
    __ __ % anthracite coal   __ __ % oil/sludge
    __ __ % western   __ __ % other (specify)
             low sulfur coal
    __ __ % lignite

    a. Source of supply? _____ (identify supplier(s) and seam(s), if known) _____

    b. What is the "proximate analysis" of each fuel if known?

       _____ % Volatile   _____ % Sulfur
       _____ % Moisture   _____ % Ash

    c. What is the chemical analysis of each ash if known?

       _____ Lithium Oxide   _____ Alumina Oxide
              (Li_2O)   (Al_2O_3)
       _____ Sodium Oxide   _____ Silica Oxide
              (Na_2O)   (SiO_2)
       _____ Potassium Oxide _____ Titanium Oxide
              (K_2O)   (TiO_2)
       _____ Magnesium Oxide _____ Phosphorus
              (MgO)   Pentoxide
                                     (P_2O_2)

       _____ Calcium Oxide   _____ Sulfur Trioxide
              (CaO)   (SO_3)
       _____ Iron Oxide   _____ Percent Carbon
              (Fe_2O_3)   at operating
                                     temperature

    d. What is the heating value for each fuel?_____ BTU/hr

12. What is the secondary fuel (if different from primary) and expected percent usage?

    _____ % bituminous coal   _____ % natural gas
    _____ % anthracite coal   _____ % bark/hog
                                      fuel/wood
    _____ % western low   _____ % bagasse
            sulfur coal
    _____ % lignite   _____ % black liquor
    _____ % oil (#2/#6)   _____ % municipal waste
                                      (describe)

13. Fuel consumption rate: _____tons/hr_____ or BTU/hr for primary and secondary fuels.

14. Firing rate: _____; % excess air _____

15. Flue gas characteristics:
    Flow rate _____ ACFM or SCFM
    Exit temperature _____ °F/°C
    Particulate loading _____ gr/scfd or gr/acf
    Dew point _____ °F
    SO_2 _____ lb/hr _____ PPM _____ %

16. Particulate size distribution
    At inlet to required new emission control equipment

Micron size	Percent of total particulate
_____	_____
_____	_____
_____	_____
_____	_____

17. Density or specific gravity of fly ash _____ lb ft^3

18. Fan capacity, if any, available for new system_____ in. W.G.

19. Available space for new systems:
    height _____ ft.
    length _____ ft.
    width _____ ft.

20. Description of existing emission control equipment, if any _____
    Will it remain in service? _____

21. Available plant services.
    Main line power:_____ volt _____ phase _____ frequency (Hz)
    Control system power:_____ volt _____ phase _____ frequency (Hz)
    Other power: _____ volt _____ phase _____ frequency (Hz)
    Steam:_____ lb/hr _____ °F _____ PSA
    Air: _____ SCFM _____ PSIA
    Water: _____ GPM _____ PSIA

22. Operating information (if known):
    Electricity costs ($/KWH)_____ (user cost or cost of generation)
    Operating labor rate ($/hr)_____ (direct labor)
    Maintenance labor rate ($/hr)_____
    Equipment accounting life (years)_____

23. Other comments _____
    _____
    _____

Please attach sketch, or use back of this question-
naire. Please also advise of any and all known
obstacles and/or special conditions.

Figure 14.   Industrial boiler emission control preliminary equipment selection questionnaire (Source: Research-Cottrell).

	Process					
Variable	Single Alkali	Limestone	Lime	Double Alkali	Citrate	Magnesia
Extent of Use	1	1	1	2	3	3
Percent SO_2 Removal	1	3	1	1	2	1
Space Requirements	1	2	2	3	3	3
Process Reliability	1	2	2	2	3	3
Process Simplicity	1	2	2	3	3	3
Process Flexibility	2	1	1	2	3	3
Waste Options	3	1	1	1	2	3
Investment Costs	1	2	2	2	3	3
Operating Costs	3	2	2	2	1	1
Life-Cycle Costs	2	2	2	2	2	2

1. Better than average
2. Average process characteristics
3. Worse than average

Figure 15. Selection matrix for chemical process.

single-alkali, lime, and limestone processes are the most widely used and best understood.

Percent SO$_2$ Removal

A key factor in the selection of a chemical process is the sulfur removal needed for the flue gas to meet environmental requirements. If high removal levels are required, it is desirable to use reagent systems which are chemically very active in reacting with the SO$_2$. Generally, if the reactants are relatively soluble, the reactions proceed at high rates before solubility problems limit the extent of SO$_2$ removal. The single-alkali, double-alkali, lime, and magnesia processes are most effective in high SO$_2$ removal because of the high levels of dissolved reactive alkali inherent in these systems.

Space Requirements

At military installations in the plains States, space may be no problem; however, in major urban centers, retrofitting a scrubber on an existing power plant may limit the choices considerably. The single-alkali process is most convenient in this regard; the double-alkali, citrate, and magnesia processes require a lot of space, and present even more problems if regeneration is also to be done at the power plant.

Process Reliability

A system must be reliable if it is to limit emissions properly. Reliability of FGD equipment has been a major problem in most systems installed by large utilities and is often the subject of public debate and legal controversy. The single-alkali process is probably the most reliable; but CERL's site visits have indicated that actual performance at many installations is considerably less than has been reported. The reliability of the regenerable processes, citrate and magnesia, has been ranked poorly in Figure 15 because the systems have yet to be tested in long-term operation.

Process Simplicity

At some remote military installations, personnel could die without heat. Therefore, if the FGD equipment fails, repair required to put the control system back in operating condition should be easy enough to be handled quickly by any available personnel. On the other hand, if the equipment is to be installed at a Government-owned facility with technically sophisticated personnel, simplicity may not be of prime concern. The single-alkali system involves the fewest operations, while the lime and limestone are next in level of chemical sophistication.

Process Flexibility

A flexible process is one whose design or operation can be readily changed after the system has been installed. Flexibility would be useful if, for example, the price of lime were to increase markedly in the future, and an installation wanted to shift quickly to a less expensive reagent. As the process chemistry becomes more complicated, flexibility is lost; the single-alkali process is the simplest available.

Waste Options

The single-alkali process produces a soluble product which can only be discharged with adequate permits from appropriate local regulatory authorities at various locations in the United States. The magnesia process produces a soluble intermediate product, either magnesium sulfite or sulfate, which must be regenerated to sulfur dioxide. Essentially, a full range of disposal options -- from liquid to semisolid wastes -- are available to the lime, limestone, and double-alkali processes.

Investment Costs

First costs may be important, but may not be the critical factor in selecting an FGD process, and often can be misleading (see Cost Information). An inappropriately designed process which permits low bids may be a very poor investment if the system cannot control SO_x emissions. Moreover, lowest first-cost processes tend to be balanced by higher operating costs, which are often associated with difficulties in a system. Therefore, operating costs can reflect problems in the areas discussed above -- e.g., waste disposal, process reliability, and percent of SO_2 removal. In addition, costs tend to average out over all processes; thus, the decision on the most appropriate chemical process generally is not strongly influenced by quoted or estimated costs, although these must be considered.

If low-cost alkalis, such as natural deposits of carbonate or bicarbonate, are available (as they sometimes are in the plains States), and disposal is not a problem, the single-alkali process can be quite economical.

Wet and Dry Disposal

The two FGD waste disposal methods are ponding (wet) and landfilling (dry). In ponding, the sludge is transported and deposited in the final disposal site as a slurry, while in landfilling, dewatering allows the sludge to be disposed of as a moist, but essentially solid, material. To alter the physical or chemical qualities of the sludge for easier handling and more environmentally acceptable disposal, sometimes the FGD sludge is also treated before final disposal -- for example, by adding lime and/or fly ash, soil, or various commercial fixation chemicals.

Wet disposal without treatment, illustrated in Figure 16, consists of piping the FGD sludge directly to a pond some distance from the site of the plant. The sludge may or may not include fly ash, depending on how fly ash disposal is handled, but usually consists of a slurry of 5 to 20 percent solids. Very often, thickeners at the FGD plant site reduce water content before the sludge is transported to the disposal area. This alleviates water balance problems at the disposal site and minimizes both the size of the disposal pond and transportation costs. The most costly item of this disposal system is the pond, which (1) must be large enough to contain the sludges produced during the life of the plant (or for a given length of time, if staged ponding is applied) and (2) must usually be lined with a natural or synthetic material to prevent any leaching to the surrounding water bodies.

Wet disposal with treatment involves a "fixing" agent which is added to the FGD sludge before final disposal in order to strengthen and stabilize the wastes and to reduce leaching. The fixing agent could be soil, lime, fly ash, or various synthetic additives, one of which is produced from basic, glassy, blast-furnace slag. Figure 17 shows a schematic diagram of the system of FGD waste disposal with treatment.

Wet disposal systems, with or without treatment, are widely applied at present to lime and limestone FGD scrubbing systems since these systems normally produce waste which is already in a form suitable for wet disposal. In the double-alkali FGD process, sludges are dewatered to recover the soluble scrubbing agent, and a disposable filter cake is produced. For a single-alkali process, wet disposal is the only way to dispose of the liquid effluent from the FGD system. Wet disposal is applicable to systems having either a wet or dry ash-handling facility, but land availability for the pond(s) is usually the limiting factor.

Dry disposal without treatment involves dewatering of the thickened sludge from the scrubbing system by centrifuges or vacuum filters, mixing the dewatered sludge with dry fly ash, and transporting it by truck to a landfill (Figure 18). The waste at the final disposal site is usually a solid material with 10 to 40 percent moisture. Therefore dams or dikes are not required in landfilling; however, diversion and drainage facilities are usually needed for runoff rainwater.

Dry disposal with treatment is similar to the above option, except that fly ash, lime, and sometimes other additives, are mixed with the dewatered FGD sludge in a muller or pug mill or similar mixers to ensure thorough mixing. The final product, a dry solid, much the same as compacted fly ash or clayey soil, can then be trucked to a landfill for final disposal (Figure 19).

Except in the case of the double-alkali FGD process, dry disposal systems, with or without treatment, are not widely used now, probably due to their relatively high operating cost compared with wet systems. Also, dry disposal may be difficult or uneconomical for systems with insufficient dry fly ash -- e.g., plants with cyclone-fired boilers or wet ash-handling
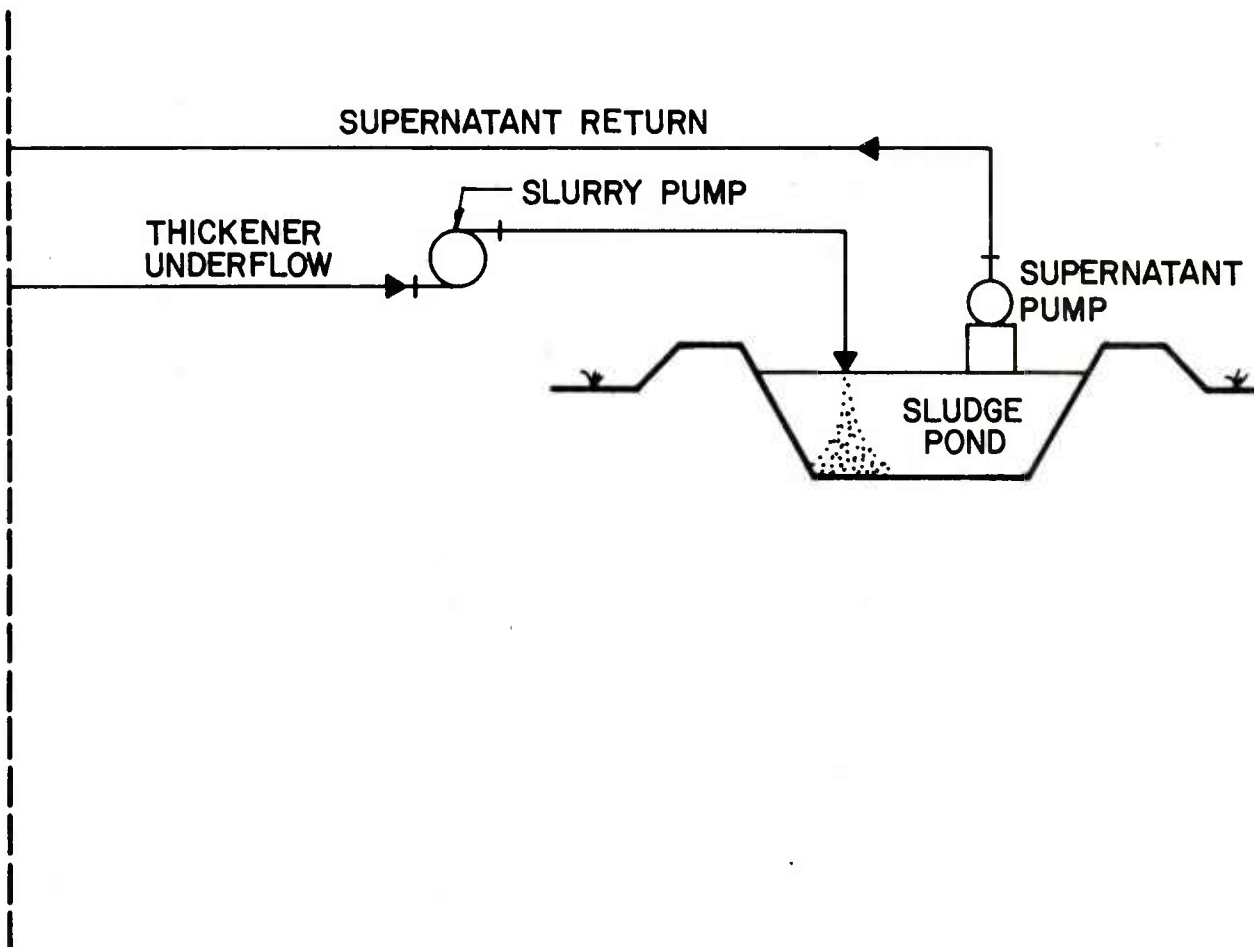
Figure 16.   Schematic diagram of wet disposal w·thout treatment.   (From
W. A. Duvel, et al., State-of-the-Art FGD Sludge Fixation,
EPRI FP-671, final report from Michael Baker, Jr., Inc., to
Electric Power Research Institute, January 1978.)
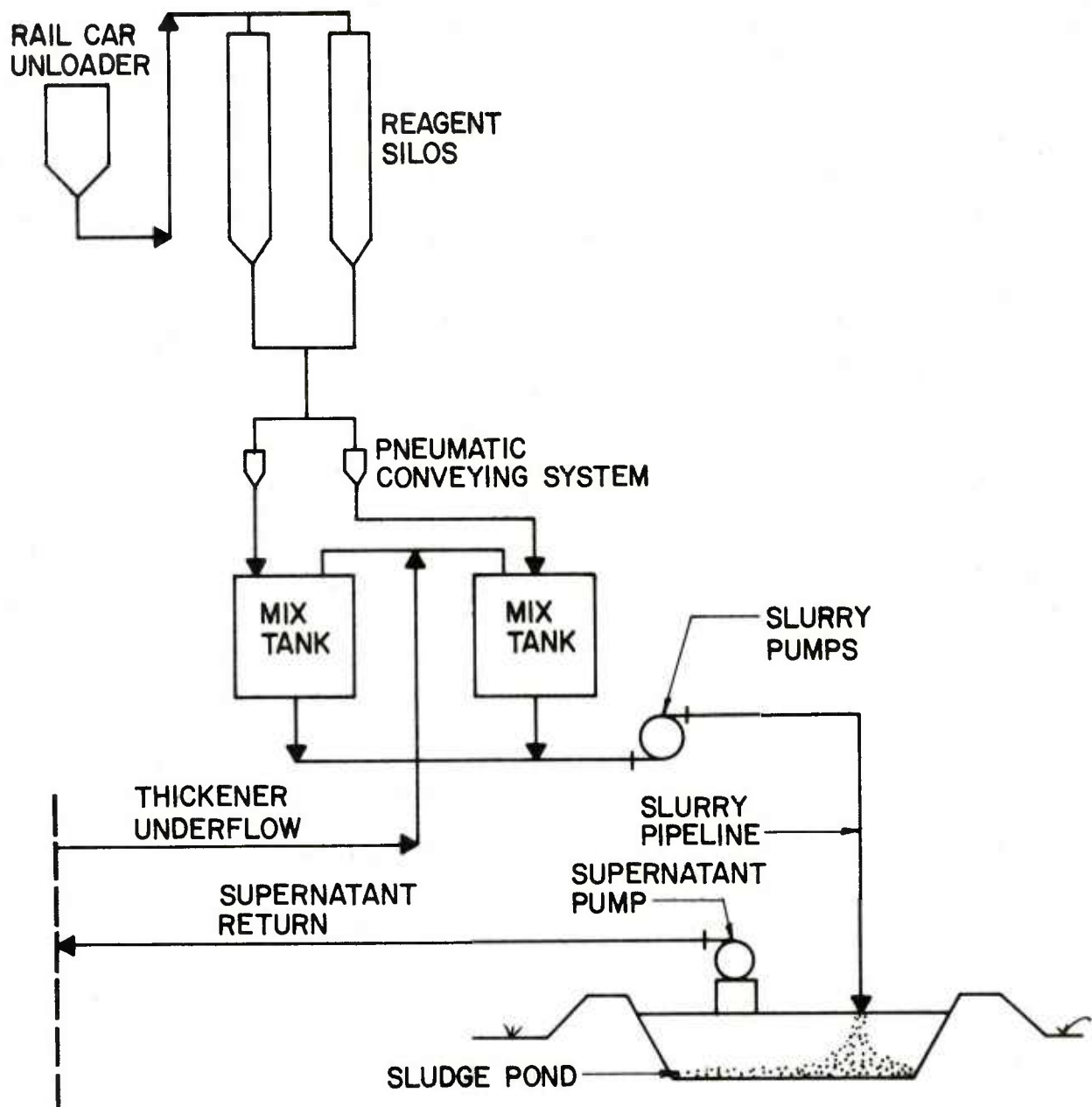
Figure 17. Schematic diagram of wet disposal with treatment. (From W. A. Duvel, et al., State-of-the-Art of FGD Sludge Fixation, EPRIU FP-671, final report from Michael Baker, Jr., Inc., to Electric Power Research Institute, January 1, 1978.)
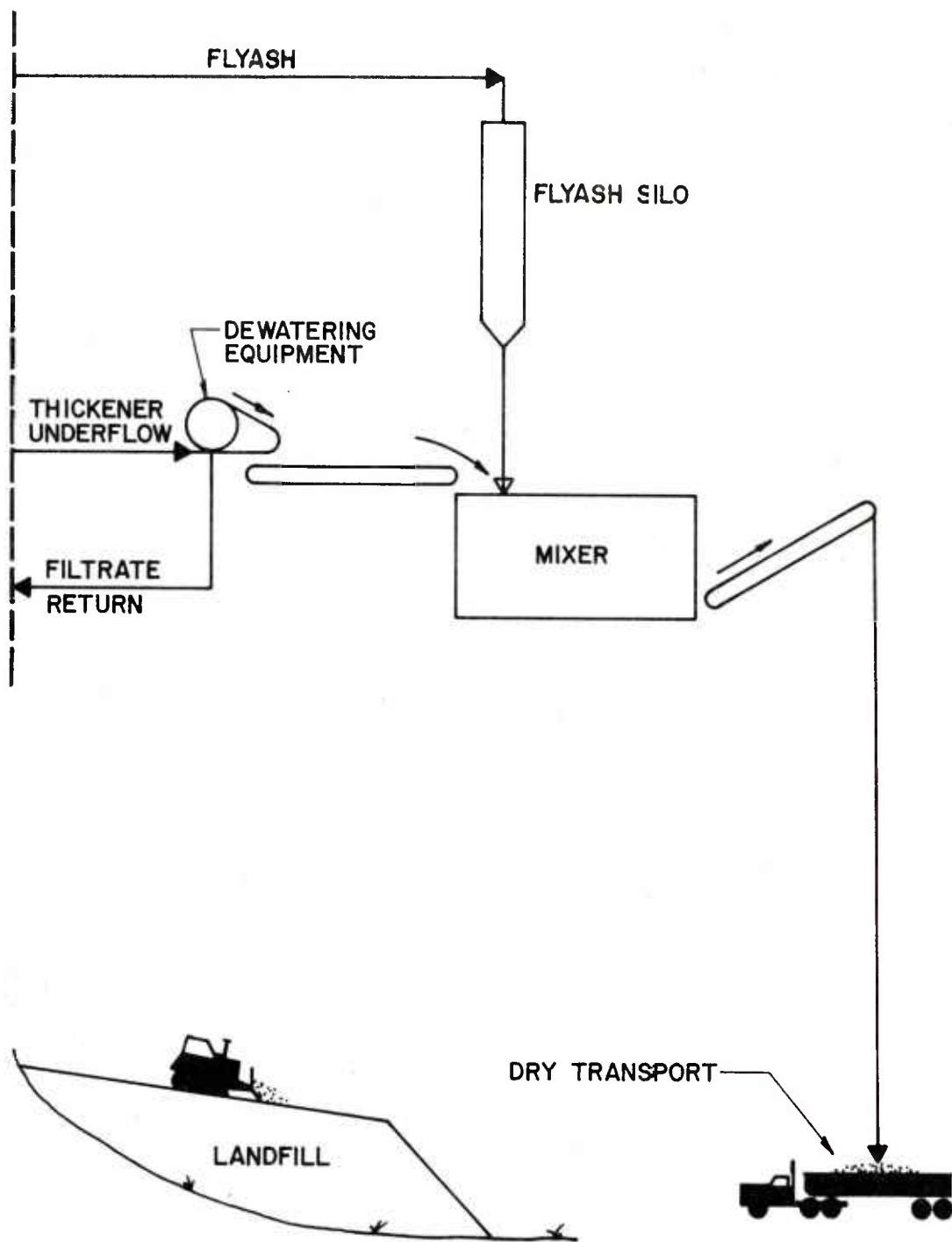
Figure 18.   Schematic diagram of dry disposal without treatment. (From
W. A. Duvel, et al., State-of-the-Art of FGD Sludge Fixation,
EPRI FP-671, final report from Michael Baker, Jr., Inc., to
Electric Power Research Institute, January 1978.)
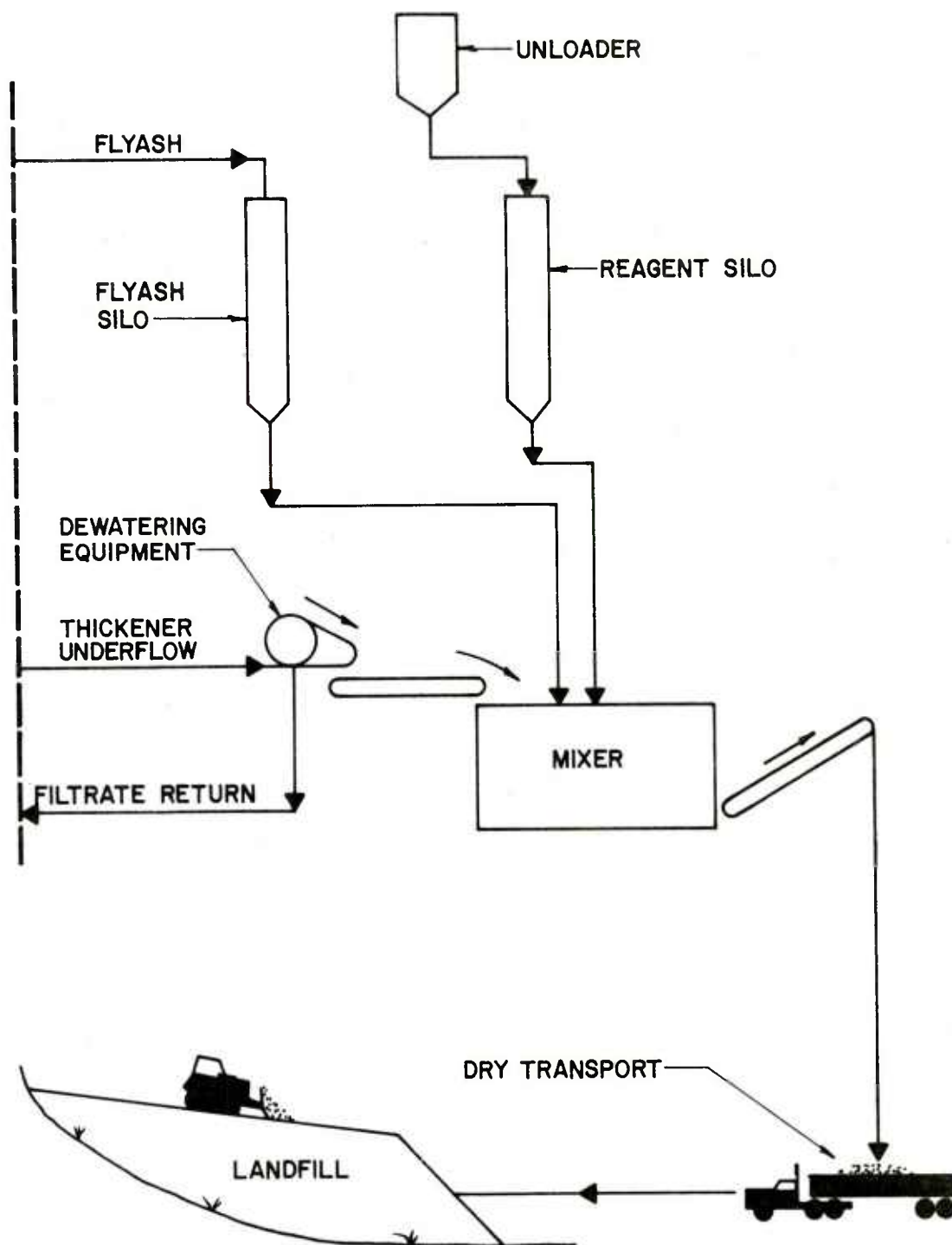
Figure 19. Schematic diagram of dry disposal with treatment. (From W. A. Duvel, et al., State-of-the-Art of FGD Sludge Fixation, EPRI FP-671, final report from Michael Baker, Jr., Inc., to Electric Power Research Institute, January 1978.)

systems, and, of course, oil-fired units. Normally, dry disposal systems are more applicable in arid regions, on systems with high ash/sludge ratios, and where land availability is a problem.

Selection of a Waste Disposal Method

In choosing an FGD waste disposal system, many influencing factors are so site specific that absolute rules for the decision-making process cannot be given. However, two principal steps normally must be taken to arrive at a decision on a disposal option. First, determine what is to be disposed of -- i.e., characterize the waste qualitatively and quantitatively. Then, in light of this information, investigate the availability of processing systems, transportation, and disposal sites.

The first step is to obtain information related to the quantity and volume of waste (or sludge) produced, as well as to its principal physical and chemical characteristics. This information will serve as an essential basis for the decision-making process to be carried out in the selection of the disposal method. Of course, the chemical composition and physical properties of the sludge cannot be accurately predicted before actual operation of the scrubber. Even within a single existing scrubbing facility, waste characteristics vary widely, depending on operating conditions. In most cases, however, the sludge composition and physical properties must be predicted for disposal planning purposes, and for compliance with the requirements of regulatory agencies. Feature requirements indicated by the Resource Conservation and Recovery Act (RCRA) and Hazardous Waste regulations will have a profound impact on waste disposal methods. One way to solve this problem is to use data from pilot plants or from actual operating installations which have a system structure similar to that of the new, planned facility -- though no two facilities are alike in all aspects.

For documentation purposes, Tables 9 and 10 compare the quality of some typical scrubber effluents with drinking water and irrigation water standards. In general, scrubber effluents do not meet drinking water standards, and frequently are too contaminated to be used for irrigation. Treatment of the sludge with chemical techniques may improve, but not necessarily solve, this disposal problem.

Another factor in the selection of a waste disposal system is the quantity and volume of waste produced. These weight and volume figures are very important in determining the amount of land area required, and in planning and designing the processing facilities and transport methods. The amount of scrubber waste produced is affected by a large number of site-specific variables (e.g., coal composition, boiler load factor, efficiencies of fly ash and SO_2 removal systems); however, tables have been compiled to permit fairly quick estimates of the weight and volume of SO_2 scrubber sludge.

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Table 9

Comparison of Constituents in FGD Sludge Liquors and
Elutriates to Irrigation Water Standards
(From W. A. Duvel, et al., FGD Sludge Disposal Manual,
EPRI FP-977, final report from Michael Baker, Jr., Inc.,
to Electric Power Research Institute, January 1979.)

Parameter, mg/L(ppm)	Long-Term Maximum Allowable Level	Short-Term Maximum Allowable Level	Range in Sludge Liquors		
Aluminum	5.0	20.0	0.03	to	0.3
Arsenic	2.0	10.0	<0.004	to	0.3
Beryllium	0.1	0.5	<0.0005	to	0.14
Boron	0.75	2.0	0.9	to	46
Cadmium	0.01	0.05	0.002	to	0.11
Chromium	0.1	1.0	0.001	to	0.5
Cobalt	0.05	5.0	<0.002	to	0.7
Copper	0.2	5.0	<0.002	to	0.6
Fluorine	2	15	0.7	to	70
Iron	5	20	0.02	to	8.1
Lead	5	10	0.0014	to	0.55
Manganese	0.2	10	0.007	to	9.0
Molybdenum	0.01	0.05	0.07	to	6.2
Nickel	0.2	2	0.005	to	1.5
Selenium	0.02	--	<0.001	to	2.2

Table 10

Comparison of Constituents in FGD Sludge Liquors and
Elutriates to Drinking Water Standards
(From W. A. Duvel, et al., FGD Sludge Disposal Manual,
EPRI FP-977, final report from Michael Baker, Jr., Inc.,
to Electric Power Research Institute, January 1979.)

| Parameter	Maximum Allowable Level	Range in Sludge Liquors		
Interim Primary Drinking Water Standards*				
Arsenic, mg/L (ppm)	0.05	<0.004	to	< 0.3
Cadmium, mg/L (ppm)	0.01	0.002	to	0.11
Chromium (VI), mg/L (ppm)	0.05	0.001	to	0.5
Fluoride, mg/L (ppm)	1.4 to 2.4	0.7		70
Lead, mg/L (ppm)	0.005	0.0014	to	0.55
Mercury, mg/L (ppm)	0.002	0.0004	to	0.07
Selenium, mg/L (ppm)	0.01	<0.001	to	2.2
Silver, mg/L (ppm)	0.05	0.005	to	0.6
Turbidity, TU	1	<3	to	10
Proposed Secondary Drinking Water Standards**				
Chloride, mg/L (ppm)	250	470	to	43 000
Copper, mg/L (ppm)	1	0.002	to	0.6
Iron, mg/L (ppm)	0.3	0.007	to	8.1
Manganese, mg/L (ppm)	0.05	0.007	to	9.0
Sulfate, mg/L (ppm)	250	720	to	30 000
Total Dissolved Solids, mg/L (ppm)	500	3200	to	95 000
Zinc, mg/L (ppm)	5	0.01	to	27
pH	6.5 to 8.5	2.8	to	12.8

*
Other parameters considered in the primary drinking water standards
besides those listed include: barium, nitrate, endrin, lindane,
methoxychlor, toxaphene, 2,4-D, 2,4,5-TP, Silvex, radium, gross alpha,
gross beta, and coliform bacteria. These parameters are either not
relevant to FGD sludges or no data are available at present.
**
Other parameters considered in the proposed secondary drinking water
standards besides those listed include: foaming agents, hydrogen
sulfide, color, corrosivity, and odor. These parameters are either
not relevant to FGD sludges or no data are available at present.

First, the coal consumption can be estimated from the following expression:

$$\text{coal used (tons/yr)} = 4.38^* \times \text{boiler heat input (Btu/hr)} \times \text{capacity}$$

$$\frac{\text{factor}}{\text{(coal heating value (Btu/hr)}} \qquad \text{[Eq 1]}$$

where: the capacity factor is the fractional seasonal use of the
maximum unit capacity (i.e., 0.00 to 1.00)
4.38 = conversion factor (1 lb/hr = 4.38 ton/yr).

Using this calculation of coal consumption and the designed plant life, the quantity (dry and slurry) and volume of scrubber sludge can be estimated quickly from Tables 11 and 12. These tables have been calculated based on three major assumptions:

1. No excess reagent, grit, or fly-ash quantities are included (amounts of excess reagent and grit are usually insignificant compared with the total waste; fly-ash quantity can be easily deduced from the coal consumption, coal ash content, and particulate removal efficiency).

2. SO_2 removal efficiencies are assumed to meet the proposed revised Environmental Protection Agency's (EPA) New Source Performance Standards (NSPS) for large boilers.** These efficiencies are shown as a function of coal sulfur content and coal heating value in Table 13.

3. SO_4/SO_3 ratio in the sludge, without better information, can be assumed either as 50/50 or as 80/20 for limestone scrubbers and 20/80 for lime scrubbers.

For example, Table 11 indicates that 1000 tons (907.2 MT) of a 10,000 Btu/lb (23 244 kJ/kg) coal with 1 percent sulfur content would produce 45.9 tons (41.64 MT) of dry solids having an 80/20 SO_4/SO_3 ratio. On the assumption that the waste is ponded as a 45 percent solids slurry, the other quantities, estimated with the help of Table 12, would be:

Volume = 45.9 (41.64) x 1.940 (1.483) = 89.1 cu yd (61.75 m^3)

Slurry Weight = 45.9 (41.64) x 2.250 (2.041) = 103.3 tons (84.99 MT)

---

* 8.76 = conversion factor for metrices (1 kg/hr = 28.76 MT/yr).
** A new NSPS is pending for industrial-size boilers, and will require a 70 percent reduction of potential SO_2 emissions. See John Molberg, "A Graphical Representation of the New NSPS for Sulfur Dioxide," Journal of the Air Pollution Control Association, Vol 30, No. 2 (February 1980), p 172.

Table 11

Weight of Dry Sludge Resulting from Scrubbing Flue Gas
From 1000 Tons (907.2 MT) of Coal to Meet
Proposed Federal NSPS
(From W. A. Duvel, et al., FGD Sludge Disposal Manual,
EPRI FP-977, final report from Michael Baker, Jr., Inc.,
to Electric Power Research Institute, January 1979.)

Btu/lb(kJ/kg)	SO4/SO3	Coal Sulfur %						
		0.5	1	2	3	4	5	6
		Dry Sludge Weight, tons (MT)						
7,000 (16,271)	80/20	21.9 (19.9)	45.9 (41.6)	91.8 (83.3)	138 (125.2)	184 (166.9)	235 (213.2)	285 (258.5)
	50/50	20.2 (18.3)	42.3 (38.4)	84.6 (76.7)	127 (115.2)	169 (153.3)	216 (196.0)	262 (237.7)
	20/80	18.5 (16.8)	38.7 (35.1)	77.4 (70.2)	116 (105.2)	155 (140.6)	198 (179.6)	240 (217.7)
7,500 (17,433)	80/20	21.7 (19.7)	45.9 (41.6)	91.8 (83.3)	138 (125.2)	184 (166.9)	232 (210.5)	282 (255.8)
	50/50	20.0 (18.1)	42.3 (38.4)	84.6 (76.7)	127 (115.2)	169 (153.3)	214 (194.1)	259 (235.0)
	20/80	18.3 (16.6)	38.7 (35.1)	77.4 (70.2)	116 (105.2)	155 (140.6)	196 (177.8)	237 (215.0)
8,000 (18,596)	80/20	21.4 (19.4)	45.9 (41.6)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	282 (255.8)
	50/50	19.7 (17.9)	42.3 (38.4)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	259 (235.0)
	20/80	18.1 (16.4)	38.7 (35.1)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	237 (215.0)
8,500 (19,758)	80/20	21.6 (19.6)	45.9 (41.6)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	282 (255.8)
	50/50	19.5 (17.7)	42.3 (38.4)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	259 (235.0)
	20/80	17.8 (16.1)	38.7 (35.1)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	237 (215.0)
9,000 (20,920)	80/20	20.9 (19.0)	45.9 (41.6)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	278 (252.2)
	50/50	19.3 (17.5)	42.3 (38.4)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	257 (233.2)
	20/80	17.6 (16.0)	38.7 (35.1)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	235 (213.2)
9,500 (22,082)	80/20	20.6 (18.7)	45.9 (41.6)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	275 (249.5)
	50/50	19.0 (17.2)	42.3 (38.4)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	254 (230.4)
	20/80	17.4 (15.8)	38.7 (35.1)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	232 (210.5)
10,000 (23,244)	80/20	20.4 (18.5)	45.9 (41.6)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	275 (249.5)
	50/50	18.8 (17.1)	42.3 (38.4)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	254 (230.4)
	20/80	17.2 (15.6)	38.7 (35.1)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	232 (210.5)
10,500 (24,407)	80/20	20.1 (18.2)	45.4 (41.2)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	275 (249.5)
	50/50	18.6 (16.9)	41.8 (37.9)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	254 (230.4)
	20/80	17.0 (15.4)	38.3 (34.7)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	232 (210.5)
11,000 (25,569)	80/20	20.0 (18.1)	45.4 (41.2)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	275 (249.5)
	50/50	18.3 (16.6)	41.8 (37.9)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	254 (230.4)
	20/80	16.8 (15.2)	38.3 (34.7)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	232 (210.5)
11,500 (26,731)	80/20	19.6 (17.8)	44.9 (40.7)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	275 (249.5)
	50/50	18.1 (16.4)	41.4 (37.6)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	254 (230.4)
	20/80	16.6 (15.1)	37.8 (34.3)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	232 (210.5)
12,000 (27,893)	80/20	19.4 (17.6)	44.3 (40.2)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	275 (249.5)
	50/50	17.9 (16.2)	40.9 (37.1)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	254 (230.4)
	20/80	16.3 (14.8)	37.4 (33.9)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	232 (210.5)
12,500 (29,056)	80/20	19.4 (17.6)	44.4 (40.3)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	275 (249.5)
	50/50	17.9 (16.2)	40.9 (37.1)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	254 (230.4)
	20/80	16.3 (14.7)	37.4 (33.9)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	232 (210.5)
13,000 (30,218)	80/20	18.9 (17.1)	43.9 (39.8)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	275 (249.5)
	50/50	17.4 (15.8)	40.4 (36.7)	84.6 (76.7)	127 (115.2)	169 (153.3)	212 (192.3)	254 (230.4)
	20/80	15.9 (14.4)	37.0 (33.6)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	232 (210.5)
13,500 (31,380)	80/20	18.6 (16.9)	43.9 (39.8)	91.8 (83.3)	138 (125.2)	184 (166.9)	230 (208.7)	275 (249.5)
	50/50	17.2 (15.6)	40.4 (36.7)	84.6 (76.7)	127 (115.2)	165 (149.7)	212 (192.3)	254 (230.4)
	20/80	15.7 (14.2)	37.0 (33.6)	77.4 (70.2)	116 (105.2)	155 (140.6)	194 (176.0)	232 (210.5)

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Table 12

Quantity of SO_x Sludge Slurry Containing 1 Ton (1.1 MT) of Sludge
Solids of 2.4 Specific Gravity
(From W. A. Duvel, et al., FGD Sludge Disposal Manual,
EPRI FP-977, final report from Michael Baker, Jr., Inc., to
Electric Power Research Institute, January 1979.)

Percent Solids in Sludge Slurry	Slurry Quantity/MT of Sludge Solids			
	Volume		Weight	
	cu yd (m^3)		ton (MT)	
5	23.000	(17.585)	20.000	(18.144)
10	11.200	(8.563)	10.000	(9.072)
15	7.210	(5.512)	6.670	(6.051)
20	5.230	(3.999)	5.000	(4.536)
25	4.050	(3.096)	4.000	(3.629)
30	3.250	(2.485)	3.330	(3.021)
35	2.700	(2.064)	2.860	(2.594)
40	2.270	(1.736)	2.500	(2.268)
45	1.940	(1.483)	2.250	(2.041)
50	1.680	(1.284)	2.000	(1.814)
55	1.460	(1.116)	1.820	(1.651)
60	1.280	(0.979)	1.670	(1.515)
65	1.130	(0.864)	1.540	(1.397)
70	1.000	(0.765)	1.430	(1.297)
75	0.888	(0.679)	1.330	(1.207)
80	0.789	(0.603)	1.250	(1.134)
85	0.702	(0.537)	1.180	(1.070)

Table 13

SO$_2$ Removal Efficiency Required to Meet Tentative EPA Standards
(From W. A. Duvel, et al., <u>FGD Sludge Disposal Manual</u>,
EPRI FP-977, final report from Michael Baker, Jr., Inc., to
Electric Power Research Institute, January 1979.)

%S Btu/lb (kJ/kg)	0.5	1	2	3	4	5	6
7,000 (16 296)	86					92	93
7,500 (17 460)	85					91	92
8,000 (18 624)	84						92
8,500 (19 788)	83						92
9,000 (20 952)	82						91
9,500 (22 116)	81						
10,000 (23 280)	80			90%			
10,500 (24 444)	79	89					
11,000 (25 608)	78	89					
11,500 (26 772)	77	88					
12,000 (27 936)	76	87					
12,500 (29 100)	75	87					
13,000 (30 264)	74	86					
13,500 (31 428)	73	86					

Note: These tentative standards are based on a paper "The Clean Air
Act Amendments of 1977--New Dimensions in Air Quality Manage-
ment," by David G. Hawkins, presented at the EPA FGD Symposium,
November

The tentative standards were stated as:

1. 90% removal of SO$_2$ from the flue gas.

2. A maximum allowable emission of 1.2 lb SO$_2$/10^6 Btu
(2.79 mg/kJ .

3. A maximum control limit of 0.2 lb SO$_2$/10^6 Btu
(0.47 mg/kJ).

A summary of the information which must be considered for the various waste disposal methods is shown in Figure 20. Again, a ranking of 1 indicates better-than-average expectations, 2 signifies average characteristics of the disposal method, and 3 suggests less-than-desirable performance. For each of the processes, the ratings for reliability, simplicity, and extent of use are the same. Since the reliability and complexity of the scrubbing processes are sufficient problems in themselves, the added difficulties of unfavorable reliability and complexity which might be associated with some waste disposal options should be avoided.

Wet disposal without treatment is the most widely used disposal method in the United States, and chemical treatment is used least. Wet disposal is also the most flexible of the disposal options, because no matter what changes might be made in the scrubbing process in the future, it is difficult to conceive of any waste product which could not be piped or trucked to a suitably lined pond. Even elemental sulfur might be ponded if necessary. Dry disposal with treatment takes the least land for waste storage, while wet ponding requires the most. When health, safety, and aesthetics are considered, dry disposal with treatment is probably the most acceptable sludge disposal method.

The chemical treatment techniques for wet and dry disposal are most limited with respect to future scrubbing options because the chemical recipe to be employed in treatment is sensitive to the composition and form of the raw scrubber waste. If a different process is used in the future, or a different by-product is produced, the appropriate formula to chemically fix this new by-product may not be the same as before.

In contrast to the costs of the chemical scrubbing processes, the costs of waste disposal do not seem to be roughly equal for all disposal methods -- especially when the expense of maintaining the disposal site indefinitely is considered. Based on information available now, the chemical stabilization treatments seem to be somewhat more expensive than disposal without treatment, but there is no low-cost option for which all probable environmental problems have been solved.

In summary, the availability of geologically acceptable land for waste storage and the existence of alternative commercial uses for the various wastes will be critical factors in selecting the most appropriate disposal method. In the long run, waste disposal is still the most troublesome problem facing the use of FGD equipment, and should be given at least as much careful attention as the selection of the chemical process to be employed in the scrubber itself.

Variable	Wet Disposal without treatment	Wet Disposal with treatment	Dry Disposal without treatment	Dry Disposal with treatment
Extent of Use	1	3	2	3
Space Requirements	3	2	2	1
Process Reliability	1	3	2	3
Process Simplicity	1	3	2	3
Environmental Impact	3	2	2	1
Process Options	1	3	2	3
Investment Costs	1	3	1	2
Operating Costs	1	2	2	3
Life Cycle Costs	1	3	2	3

1. Better than average
2. Average disposal characteristics
3. Worse than average

Figure 20. Selection matrix for sludge disposal.

Cost Information

One of the most important aspects of sulfur control technology is the cost involved in its use. When a specific control process is being selected, estimates of both capital and operating costs should be as accurate as possible. Unfortunately, in an emerging technology such as FGD, it is difficult to estimate costs reliably.

This discussion of costs has therefore been prepared to assist in planning and decision-making. Cost information has been developed from a number of sources and generalized as a function of plant size, coal sulfur content, and method of waste disposal for the six FGDs studied. The level of sulfur removal is not reported as a variable. Except at extremely high sulfur-removal requirements, the increased first costs for added sulfur removal are outside the accuracy of these cost estimates. The accuracy of the estimates included in this chapter is probably about $\pm$ 25 percent for new facilities. Retrofit cost requirements for some older or space-limited facilities may be a factor of two higher than these estimates. Finally, the actual system reliability desired by DOD may require control equipment in addition to that being costed at present.

Perhaps as FGD systems are used more, their costs will decrease. But this chapter summarizes CERL's best estimates for the purchase prices and operating costs of technology currently being supplied and used by industry.

Basis of Information and Calculations

This chapter provides data on installed capital costs, and annually or cyclically recurring costs, for six new and retrofit FGD systems. These processes are applied to coal-fired boilers having nominal capacities of 25, 50, 100, 150, and 250 x 10^6 Btu/hr (7.33, 14.65, 29.31, 43.96, and 73.27 MW) input, and burning fuel of 0.4 and 4.5 percent sulfite. The information was compiled from two major sources: vendors and owners of FGD systems, and a computerized model.

Vendors and Owners of FGD Systems

Actual costs reported by current users and estimates provided by vendors of different FGD systems were gathered from information submitted by letter, site visits, or telephone interviews. For comparison purposes, these data were then adjusted to October 1978 tests, which correspond to an annual escalation rate of about 7.5 percent. No other adjustments were attempted, and the cost figures for each FGD system therefore include all cost contributions for other relevant features (e.g., origin, size and type, load factor, space availability) which result from site-specific factors.

Computerized Model

Battelle's Columbus Laboratories developed a model -- based on existing process designs and cost estimate data -- to predict the capital costs, annualized costs, and energy requirements for FGD systems large enough to be of interest to electric utilities.[4] These steam-electric plants are generally at least an order of magnitude larger than power plants used at military installations. Therefore, a scaling procedure was developed from the accumulated data bank, which contains information on equipment and costs, as well as labor, material, and utility requirements for each FGD process. This procedure allowed costs to be projected for smaller plants.

For instance, flue gas flow rate and amount of sulfur scrubbed are two major variables on which the scaling procedure was performed to calculate cost estimates for the five boiler sizes (25 to 250 x 10^6 Btu/hr [7.33 to 73.27 MW]) of interest to the Army. The computerized model first calculated the flue gas flow rate and the level of sulfur scrubbing based on a set of process-related and site-specific assumptions (e.g., coal characteristics, boiler size), and then with the scaling procedure determined the installation cost of equipment and materials, and utilities requirements. Capital and annualized costs were then estimated by Guthrie's method, using a set of economics-related assumptions (e.g., cost adjustment factors, unit costs of materials, utilities).[5]

To give an estimated range of the capital costs and cyclically recurring cost elements for each FGD process at a particular boiler size, two coals were selected for study: one of 0.4 percent sulfur, 8,000 Btu/lb (18 624 kJ/kg) (low-sulfur coal) and the other 4.5 percent sulfur, 12,000 Btu/lb (27 936 kJ/kg) (high-sulfur coal). Two methods of waste disposal -- ponding (wet) and landfill (dry) -- were also investigated when applicable to any of the six FGD processes studied. The essential basis and assumptions which the model employed in this task of cost estimation are summarized in Table 14. It has also been assumed either that the equipment can be located outdoors without extensive protection needed to prevent freeze-ups, or that a building is already available.

Capital Costs for FGD Equipment

The actual capital costs of currently installed FGD systems, as reported by users or estimated by vendors, are presented in Table 15. A cost index was used to convert all these costs to an equivalent time basis (circa October 1978) for comparison purposes. Costs tend to average at about $1400/$10^6$ Btu/hr ($4777/MW).

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[4] Battelle's Columbus Laboratories, Modeling of Economic and Energy Impacts of Flue Gas Desulfurization Systems, In-house Study (September 1978).

[5] K. M. Guthrie, "Capital Cost Estimating," Chemical Engineering (March 24, 1969), pp 114-142.

Table 14

Basis and Assumptions Used by the Computerized Model in the
Cost Estimation of Different FGD Systems

A. **FGD Processes Studied**

1. Wet Limestone
2. Lime Scrubbing
3. Single Alkali
4. Double Alkali
5. Magnesium Oxide
6. Citrate

B. **Nominal Boiler Sizes Studied**, million Btu/hr (GJ/hr) input

25, 50, 100, 150, 250   (Assuming a conversion efficiency of
(26.4, 52.8, 105.5, 158.3,   36 percent, these sizes correspond
263.8)   to plant sizes of 2.6, 5.3, 10.6,
                               15.8, and 26.4 MWe, respectively.)

C. **Assumptions for Combustion/Humidification Calculations**

Coal Characteristics:	0.4 percent S and 8,000 Btu/lb (18 624 kJ/kg) and 4.5 percent S and 12,000 Btu/lb (27 936 kJ/kg)
Excess Air	70 percent over stoichiometric requirement
Air Leakage	15 percent of flue gas
Air Humidity	0.002 lb H_2)/lb (.002 kg/kg) dry air
Ash Overhead	30 percent of total ash
Standard Temperature	68°F (20°C)

D. **Design Variables**

SO_2 Removal	85 percent for high-sulfur coal 0.5 lb $SO_2/10^6$ Btu (0.216 mg/kJ) SO_2 emission for low-sulfur coal (equivalent to about 50 percent SO_2 removal)
Flue Gas Temperature	350°F (176.7°C) at scrubber inlet, adiabatic at scrubber outlet (about 115°F to 125°F [46.1 °C to 51.7°C]).
No Flue Gas Reheat	
No Equipment Redundancy	
Waste Disposal	35 percent solid-content sludge (if wet, 1.0 mile (1.609 km) of pipeline)

Table 14 (Cont'd)

E. Economic Data and Assumptions

Cost Index (Marshall &
  Swift)   561.8 (October 1978)
Maintenance-Related Charges   5 percent of installation costs *
No Land Cost
No Capital-Related Charges   (i.e., no DCF rate, no tax, no
                                  insurance, etc.)
Operating Labor $10/man-hour (includes overhead)
                                Irrespective of boiler size, assign
                                  8760 man-hr/yr to all FGD systems,
                                  except Magnesium Oxide and Citrate
                                  processes, for which 13,140
                                  man-hr/yr is used.

F1. Adjustment Factors for Capital Costs Calculation

Foundation   7.2 percent of total equipment cost
Mechanical   8.5 percent of total equipment cost
Electrical   6.0 percent of total equipment cost
Instrumentation   3.0 percent of total equipment cost
Distributables   3.0 percent of total equipment cost
Miscellaneous   2.0 percent of total equipment cost
Contingency   5.0 percent of installation cost
Supervision and Administra-
  tion   5.0 percent of installation cost
Contractor Fee   3.0 percent of installation cost

F2. Raw Material Costs (applicability dependent on FGD process)

Citric Acid $ 0.70/lb ($ 1.54/kg)
Coke $ 60.00/ton($ 65.14/MT)
Lime $ 28.00/ton($ 30.86/MT)
Limestone $ 7.30/ton($ 8.05/MT)
Magnesium Oxide $150.00/ton($165.34/MT)
Hydrogen Sulfide $ 62.60/ton($ 69.00/MT)
Soda Ash $ 98.00/ton($108.02/MT)

F3. Utility Costs (applicable dependent on FGD processes)

Process Water $1.00 /1000 gal ($ .264/1000L)
Cooling Water $0.25 /1000 gal ($ .066/1000L)
Electricity $0.035/kWh
Steam $4.00 /1000 lb ($8.818/1000kg)
Purge Stream Treatment $0.15 /1000 gal ($0.040/1000L)

F4. Fuel Costs

Fuel Oil $0.60 /gal ($0.159/L)

Table 14 (Cont'd)

F5. <u>By-Product Credits</u>

Sulfur	$40.00/ton ($44.10/MT)
Sulfuric Acid	$45.00/ton ($49.60/MT)

*Installation Cost = Total equipment cost
+ Foundation surcharge
+ Mechanical surcharge
+ Electrical surcharge
+ Instrumentation surcharge
+ Distributables surcharge
+ Miscellaneous surcharge
+ Cost of pond and pipeline (if any).

Table 15

Capital Costs of Different FGD Systems as Provided by Users and/or Vendors

Unit No.	Unit I.D.	FGD Process	Boiler Size, 10⁶ Btu/hr (GJ/hr)	% S in Coal Used	% SO₂ Removal	Status	Flue Gas Reheat	Capital Cost, $ million Given	Adjusted to 10/1978
1.	Rickenbacker AFB (Lockbourne, OH)	Lime/Limestone	210 (222)	3.5	70-85	Retrofit	No	2.20	2.86
2.	GM Chevrolet Plant (Parma, OH)	Dilute Double Alkali	300 (317)	2-3	80	Retrofit	No	3.20	3.92
3.	Caterpillar Tractor (Joliet, IL)	Dilute Double Alkali	215 (227)	2-4	85-90	Retrofit	No	4.0	4.0
4.	Armco Steelworks (Middletown, OH)	Lime	370 (390)	0.6-1.0	Not measured	Retrofit	No	1.25	1.58
5.	ITT-Rayonier Cellulose Plant (Fernandina Beach, FL)	Single Alkali	530 (559)	2.75	75	Retrofit	No	0.8	1.31
6.	Kerr-McGee Chemical Corp. (Trona, CA)	Single Alkali	1200 (1266)	1.5	96	New	Steam	6.0	6.0
7.	Great Western Sugar Co. (Fremont, OH)	Single Alkali	150 (158)	0.6-1.0	20-25	New	No	0.8	1.0

The computerized model provided ranges of estimated capital costs of six FGD systems (newly installed), for the five boiler sizes and for coals of sulfur contents from 0.5 to 4.5 percent (Table 16). It should be noted that, as calculated, these estimated capital costs include the equipment installation expenses, plus surcharges (foundation, mechanical, electrical, instrumentation, distributables, miscellaneous), costs for supervision/administration and contingencies, and modest fees for return on investment by the supplier. The assumed magnitudes of these cost adjustment factors are shown in Table 14. Graphical display of these capital cost estimates is shown in Figures 21 and 22.

As expected, lowest capital costs are associated with nonregenerative scrubbers. If disposal of the liquid wastes from single-alkali systems without ponding is feasible at a specific facility, single-alkali technology offers further savings in first costs. Single-alkali technology also tends to be more reliable. As will be shown later, however, operating costs for single-alkali systems, particularly for higher sulfur coals, are somewhat greater than those for scrubbing systems that produce solid wastes or solid by-products. The lime and limestone processes are essentially identical in first costs; in fact, a scrubber at Rickenbacker Air Force Base, OH, can be used with either lime or limestone. The mechanical equipment for a double-alkali scrubber can be significantly more expensive, but because it produces a relatively water-free solid waste, combined equipment and waste-disposal costs are similar to other throwaway processes. If a filtration system for slurry wastes is specified for future lime and limestone systems, first costs for such technology might be the most attractive overall investment at many locations. The citrate and magnesium oxide processes are the most expensive. They provide the most acceptable products from the viewpoint of the environment as a whole, but many installations may not be able to dispose of these by-products -- i.e., sulfur and H_2SO_4 -- by selling them to commercial markets. Disposal of by-product H_2SO_4 or elemental sulfur under contract to an H_2SO_4 manufacturer is a method of integrated operation that has yet to be proven practical. So far, even the electric utility industry has not successfully marketed sulfur -- after trying for a decade. Nonetheless, for installations such as ammunition plants and large military bases, this technology may be of considerable interest.

Operating Costs for FGD Equipment

Cyclically recurring costs of different installed FGD systems as reported by users or vendors are shown in Table 17. Since most of these units are still in the demonstration phase of commercialization, it is difficult to determine a firm basis on which annual figures can be estimated. Thus, only the rate of materials required or produced is adjusted to the basis of a 7008 hr/yr operation. Labor, maintenance and repair, and operating costs, whenever available, are reported as they were quoted by users. The year during which these costs were reported is also indicated in the table.

Table 16

Estimated Capital Costs of New FGD Systems
(Costs are shown in million dollars)

Boiler Size 10^6 Btu/hr (GJ/hr)	25 (26.4)	50 (52.8)	100 (105.5)	150 (158.3)	250 (263.8)
Single Alkali	0.5-0.8	0.8-1.2	1.2-2.0	1.7-2.7	2.5-3.9
Wet Limestone	0.6-1.1	0.9-1.5	1.3-2.2	1.8-2.8	2.6-3.8
Lime	0.6-1.1	0.9-1.5	1.3-2.2	1.8-2.8	2.6-3.8
Double Alkali	0.5-1.0	0.8-1.5	1.2-2.3	1.6-3.0	2.2-4.2
Magnesium Oxide	0.8-1.6	1.1-2.1	1.6-3.0	2.0-3.7	2.7-4.7
Citrate	0.6-1.4	0.9-1.9	1.3-2.7	1.6-3.3	2.2-4.3

Note: 1. Costs shown are estimated at 10/1978 (Marshall & Swift index = 561.8).

2. No redundancy was included.

3. Estimated ranges are applicable to coals of 0.4 percent to 4.5 percent sulfur and of 8,000 to 12,000 Btu/lb (8440 to 12 660 GJ/hr) heating value.

4. Retrofit factor = 1.0 to 1.8, dependent upon the size and age of the boiler. The smaller the size and the older the age, the closer this factor approaches 1.8. This factor should be used to multiply the above estimated capital costs when retrofitting an FGD system to an existing plant.

5. Practically, MgO and citrate FGD processes are not applicable and not recommended for the above boiler sizes, unless an FGD central treatment facility is used commonly for a multiplicity of such small boilers.
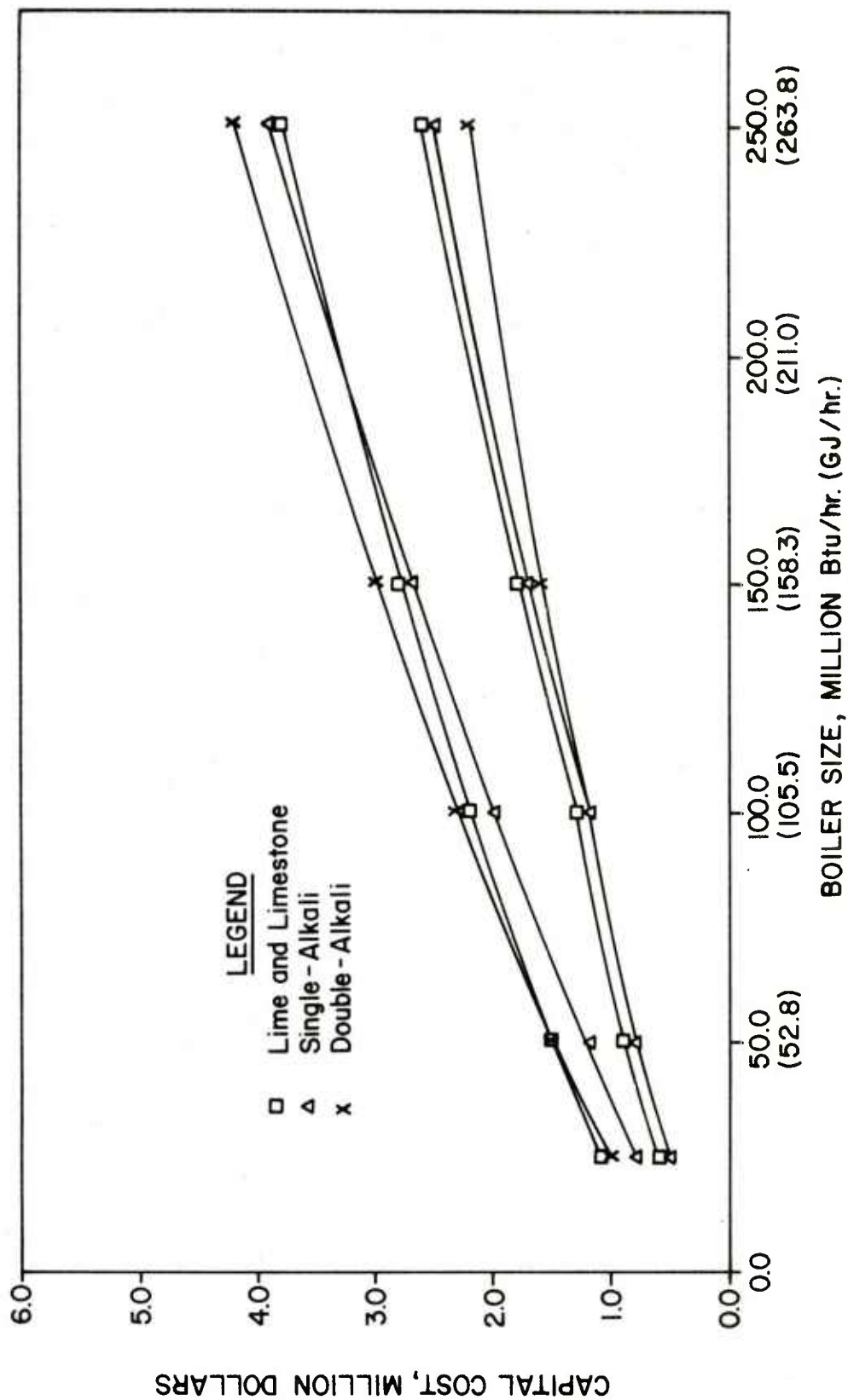
87

Figure 21. Estimated capital costs of new FGD systems: lime, limestone, single-alkali, double-alkali (ranges shown as high and low limits).
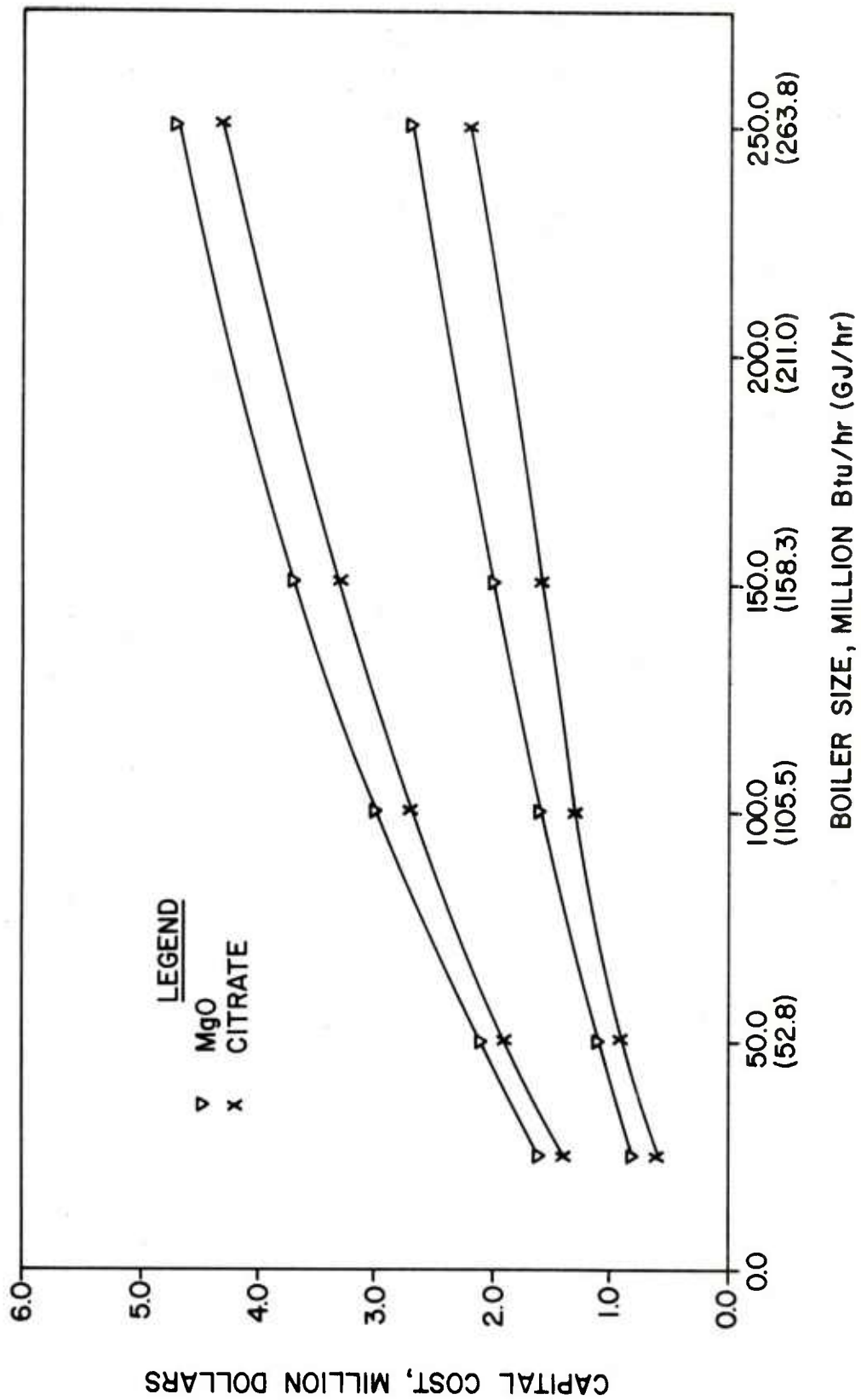
88

Figure 22. Estimated capital costs of new FGD systems: MgO, citrate (ranges shown as high and low limits).

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Table 17

Cyclically Recurring Cost Elements of Actual FGD Systems as Reported
by Users and Vendors

Unit No.		Boiler Size, 10⁶ Btu/hr (GJ/hr)	Electricity (10⁶kWh/yr)	Water, 10³Kgal/yr (10⁶ L/yr)	Raw Materials Kton/yr (Kt/yr)	Liquid Waste 10³Kgal/yr (10⁶ L/yr)	Solid Waste Kton/yr (Kt/yr)	Operating Cost $/ton coal(yr) ($/MT)
1.	Lime/Limestone	210 (222)	3.7	10 (37.85)	3.5(3.2)[1] or 7 (6.35)[2]	6.5(24.6)	18-21 (16.3-19.1)	6.0 (6.6) (1977)
2.	Dilute Double Alkali	300 (317)	3.8	22.5(85.17)	4.3(3.0)[1] 0.8(0.7)[3]	Negligible	19-23 (17.2-20.9)	13.3(14.7) (1978)
3.	Dilute Double Alkali	215 (227)	1.9	13.5(51.10)	4.7(3.4)[1] 1.0(.9)[3]	Negligible	18.3 (16.6)	11.3(12.5) (1978)
4.	Lime	370 (390)	6.2	400(1514.16)	1.3(1.2)[1]	673(2547.6)[5]	--	No data
5.	Single Alkali	530 (559)	15	32.4(122.65)	642 Kgal/yr caustic (2.43·10⁶L/yr)	700 (2649.8)	Negligible	Not Applicable
6.	Single Alkali	1200(1266)	15	390(1476.31)	71.5 Kgal/yr (27.1·10⁶L/yr)[4]	385 (1457.4)	Negligible	Not Applicable
7.	Single Alkali	150(158)	1.3	147(556.40)	100 Kgal/yr (50% caustic) (3.79·10⁶L/yr)	147 (556.5)[5]	Negligible	Not Applicable

Note: All annual figures are average estimates on a basis of 7008 hr/yr. Unit numbers match with those shown on Table 15.

1. Lime.
2. Limestone.
3. Soda ash.
4. Burkeite solution (29 percent solids).
5. Solids content unknown.

For each of five boiler sizes, the estimated annual costs of the six FGD processes are summarized in Tables 18 through 23. The relatively wide ranges of cost reflected by these estimates indicate the effect of the sulfur content and heating value of the two coals selected for this study (one was 0.4 percent sulfur, 8,000 Btu/lb [18 624 kJ/kg]; the other 4.5 percent sulfur, 12,000 Btu/lb [27 936 kJ/kg]). Figures 23 and 24 graph these operating cost estimates.

The estimates and reported values show that for low-sulfur coal and for small boilers, the nonregenerable processes are the least expensive annually, and the four throwaway processes under consideration are roughly equivalent in operating cost. However, as the sulfur content increases, and as the costs of the added reagent increase, the single-alkali process becomes more costly. Therefore, it appears prudent to use a less expensive source of alkali, such as lime, or even limestone if the desired degree of SO_2 removal can be obtained without prohibitive excesses of limestone. The estimated costs vary widely, from about $5/ton ($5.51/MT) of coal burned to as high as $30/ton ($33.07 MT) for the smaller boiler sizes. The costs of such control technology can approach or exceed the cost of the fuel being burned; furthermore, for proper operation, the FGD system may require more skilled, and more highly paid personnel than do the furnaces or boilers currently used at various Army installations.

Waste Disposal Expenses

The costs of waste disposal are worthy of detailed discussion. Wastes produced by the FGD systems, at present, have not yet been clearly categorized by the EPA. They are classified neither as hazardous, nor as nonhazardous wastes, but rather are in a special category; the toxicity of wastes in this category is determined from case to case. Such uncertainty in waste characterization, coupled with diverse waste disposal methods and site characteristics, makes it difficult to generalize the costs of FGD waste disposal. In spite of this problem, however, a rough estimating method was used in the previous sections to approximate the capital and operating costs of the waste disposal problem associated with the six FGD systems. Of special interest is that the costs of waste disposal can significantly affect the choice of the optimum FGD system at any given location.

The most common methods of waste disposal, dry and wet, were costed, where applicable, for the six FGD processes studied. However, with the single-alkali process, the waste produced is essentially a liquid stream; so CERL considered only ponding and deep-well injection. All the other processes may use either ponding (wet) or landfill (dry) to dispose of their waste. The cost of land was not taken into account because at most military establishments this expense is not considered in accounting procedures for an individual facility on the installation.

Table 18

Annual Cyclically Recurring Cost Elements of Limestone FGD Systems*

FGD Process	Limestone				
Boiler Size, 10^6 Btu/hr (GJ/hr)	25 (26.4)	50 (52.8)	100 (105.5)	150 (158.3)	250 (263.8)
Maintenance and Repair ($)**	27-50	41-68	58-99	81-126	117-171
Labor, man-hr†	8760	8760	8760	8760	8760
Electricity (10^6 kWh)	0.4-0.8	0.9-1.7	1.8-3.4	2.6-5.0	4.4-8.5
Potable Water, 10^3 Kgal (10^6L)	0.54-1.8 (2.0-6.8)	1.1-3.6 (4.2-13.6)	2.2-7.2 (8.3-27.2)	3.3-11.0 (12.5-41.6)	5.4-18.0 (20.4-68.1)
Process Water, 10^3 Kgal (10^6L)					
Chemicals					
Limestone, Kton (kt)	0.09-1.2 (0.08-1.1)	0.18-2.3 (0.16-2.1)	0.36-4.6 (0.33-4.2)	0.54-6.9 (0.49-6.3)	0.90-11.5 (0.82-10.4)
Wastes					
Liquid, 10^3 Kgal (10^6L)	0.4-0.8 (1.5-3.0)	0.8-1.7 (3.0-6.4)	1.5-3.4 (5.7-12.9)	2.3-5.0 (8.7-18.9)	3.8-8.5 (14.4-32.2)
Solid, Kton (kt)	0.9-2.0 (0.8-1.8)	1.7-3.8 (1.5-3.4)	3.4-7.7 (3.1-7.0)	5.1-11.5 (4.6-10.4)	8.6-19.1 (7.8-17.3)
Operating Cost					
$/ton (MT) Coal	12-25 (13.2-27.6)	8-17 (8.8-18.7)	5-13 (5.5-14.3)	4-11 (4.4-12.1)	4-10 (4.4-11.0)
Cents/Btu (GJ) Input	77-105 (73.0-99.5)	48-71 (45.4-67.3)	32-53 (30.3-50.2)	27-46 (25.6-43.6)	23-41 (21.8-38.9)

*Estimated by Battelle computerized model, based on full-load boiler operation at 7008 hr/yr.

**Assuming 5 percent of installation cost (see Table 10 also).

†Assuming 24 man-hr for lime, limestone, single-alkali, and double-alkali processes.

Table 19

Annual Cyclically Recurring Cost Elements of Lime FGD Systems*

FGD Process	Lime				
Boiler Size 10^6 Btu/hr (GJ/hr)	25 (26.4)	50 (52.8)	100 (105.5)	150 (158.3)	250 (263.8)
Maintenance and Repair, ($)**	30-48	40-67	56-97	69-123	91-170
Labor, man-hr†	8760	8760	8760	8760	8760
Electricity (10^6 kWh)	0.3-0.5	0.6-1.0	1.2-2.0	1.8-3.0	3.1-5.0
Potable Water, 10^3 Kgal (10^{6}L)	2.3-4.0 (8.7-15.1)	4.6-7.0 (17.4-26.5)	9.0-14.0 (34.1-53.0)	14.0-20.0 (53.0-75.7)	23.0-34.0 (87.1-128.7)
Process Water, 10^3 Kgal (10^{6}L)					
Chemicals					
Lime, Kton (kt)	0.04-0.5 (0.04-.5)	0.09-1.1 (0.08-1.0)	0.17-2.2 (0.15-2.0)	0.25-3.2 (0.23-2.9)	0.42-5.4 (0.38-4.9)
Wastes					
Liquid, 10^3 Kgal (10^{6}L)	0.4-0.9 (1.5-3.4)	0.8-1.9 (3.0-7.2)	1.6-3.7 (6.1-14.0)	2.3-6.0 (8.7-22.7)	3.8-9.2 (14.4-3.5)
Solid, Kton (kt)	0.6-1.5 (0.5-1.4)	1.2-2.7 (1.1-2.4)	2.4-5.5 (2.2-5.0)	3.6-8.1 (3.3-7.3)	6.1-13.5 (5.5-12.2)
Operating Cost					
$/ton (MT) Coal	13-24 (14.3-26.5)	8-17 (8.8-18.7)	5-12 (5.5-13.2)	4-11 (4.4-12.1)	4-10 (4.4-11.0)
Cents/10^6 Btu (GJ) Input	79-102 (74.9-96.7)	49-69 (46.4-65.4)	33-51 (31.3-48.3)	27-44 (25.6-41.7)	22-39 (20.9-37.0)

*Estimated by Battelle computerized model, based on full load boiler operation at 7008 hr/yr.

**Assuming 5 percent of installation cost (see Table 10 also).

†Assuming 24 man-hr for lime, limestone, single-alkali, and double-alkali processes.

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Table 20

Annual Cyclically Recurring Cost Elements
of Single-Alkali FGD Systems*

FGD Process	Single Alkali				
Boiler Size 10^6 Btu/hr (GJ)	25 (26.4)	50 (52.8)	100 (105.5)	150 (158.3)	250 (263.8)
Maintenance and Repair (10^3 $)**	28-32	44-52	71-86	95-116	137-173
Labor, man-hr†	8760	8760	8760	8760	8760
Electricity, (10^6 kWh)	0.2-0.3	0.5-0.7	0.9-1.3	1.4-1.9	2.3-3.2
Potable Water, 10^3 Kgal (10^{6}L)	3-6 (11.4-22.7)	6-12 (22.7-45.4)	12-24 (45.4-90.8)	18-35 (68.1-132.5)	30-59 (113.6-223.3)
Process Water, 10^3 Kgal (10^{6}L)					
Chemicals					
Soda Ash, Kton (kt)	0.08-1.0 (0.07-0.9)	0.16-2.0 (0.15-1.8)	0.30-4.0 (0.27-3.6)	0.50-6.0 (0.45-5.4)	0.80-10.0 (0.73-9.0)
Wastes					
Liquid, 10^3 Kgal (10^{6}L)	0.4-0.9 (1.5-3.4)	0.8-1.8 (3.0-6.8)	1.5-3.6 (5.7-13.6)	2.3-5.4 (8.7-20.4)	3.9-9.0 (14.8-3.4)
Solid, Kton (kt)	0.9-2.0 (0.8-1.8)	1.7-7.6 (1.5-6.7)	3.5-8.2 (3.2-7.4)	5.2-12 (4.7-10.9)	8.7-20 (7.9-18.1)
Operating Cost					
$/ton (MT) Coal	12-33 (13.2-36.4)	8-26 (8.8-28.7)	5-23 (5.5-25.4)	5-21 (5.5-23.1)	4-20 (4.4-22.0)
Cents/10^6 Btu (GJ) Input	78-138 (73.9-130.8)	50-109 (47.3-103.3)	35-94 (33.2-89.1)	30-89 (28.4-84.4)	25-84 (23.7-79.6)

*Estimated by Battelle computerized model, based on full-load boiler operation at 7008 hr/yr.
**Assuming 5 percent of installation cost (see Table 10 also).
†Assuming 24 man-hr/day for lime, limestone, single-alkali, and double-alkali processes.

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Table 21

Annual Cyclically Recurring Cost Elements of Double-Alkali FGD Systems*

FGD Process	Double Alkali				
Boiler Size 10^6 Btu/hr (GJ/hr)	25 (26.4)	50 (52.8)	100 (105.5)	150 (158.3)	250 (263.8)
Maintenance and Repair (10^3 \$)**	23-46	35-68	54-103	70-134	97-188
Labor, man-hr†	8760	8760	8760	8760	8760
Electricity (10^6 kWh)	0.2-0.4	0.4-0.7	0.9-1.4	1.4-2.0	2.3-3.3
Potable Water, 10^3 Kgal (10^6L) ⎫ Proce-s Water, 10^3 Kgal (10^6L) ⎭	2-3 (7.6-11.4)	4-6 (15.1-22.7)	9-13 (34.1-49.2)	14-20 (53.0-75.7)	23-33 (87.1-12.5)
Chemicals					
Lime, Kton (kt)	0.04-0.5 (0.04-.5)	0.08-1.0 (0.07-.9)	0.2-2.0 (0.2-1.8)	0.23-3.0 (0.21-2.7)	0.4-5.0 (0.4-4.5)
Soda Ash, Kton (kt)	0.004-0.05 (0.004-.05)	0.01-0.1 (0.01-0.1)	0.02-0.2 (0.02-0.2)	0.02-0.3 (0.02-0.3)	0.04-0.5 (0.04-0.5)
Wastes					
Liquid, 10^3 Kgal (10^6L)	0.4-0.7 (1.5-2.6)	0.8-1.4 (3.0-5.3)	1.5-2.8 (5.7-10.6)	2.0-4.2 (7.6-15.9)	3.7-7.0 (14.0-26.5)
Solid, Kton (kt)	0.8-1.6 (0.73-1.5)	1.7-3.2 (1.5-2.9)	3.5-6.5 (3.2-5.9)	5-9.5 (4.5-8.6)	8.4-15.7 (7.6-14.2)
Operating Cost					
\$/ton (MT) Coal	12-24 (13.2-26.5)	7-16 (7.7-17.6)	5-12 (5.5-13.2)	4-11 (4.4-12.1)	4-9 (4.4-9.9)
Cents/10^6 Btu (GJ) Input	73-98 (69.2-92.9)	45-67 (42.7-63.5)	30-50 (28.4-47.4)	25-43 (23.7-40.8)	21-38 (19.9-36.0)

*Estimated by Battelle computerized model, based on full-load boiler operation at 7008 hr/yr.
**Assuming 5 percent of installation cost (see Table 10 also).
†Assuming 24 man-hr/day for lime, limestone, single-alkali and double-alkali processes.

Table 22

Annual Cyclically Recurring Cost Elements
of Magnesium Oxide FGD Systems*

Boiler Size 10⁶ Btu/hr (GJ/hr)	25 (26.4)	50 (52.8)	100 (105.5)	150 (158.3)	250 (263.8)
Maintenance and Repair.(10³ $)**	36-71	52-97	74-135	92-164	122-212
Labor, man-hr†	13140	13140	13140	13140	13140
Electricity (10⁶ kWh)	0.4-0.7	0.8-1.3	1.5-2.6	2.3-3.9	3.9-6.5
Potable Water, 10³ Kgal (10⁶L)	4.5-11.0 (17.0-41.6)	8.9-21.4 (33.7-81.0)	1.8-43.0 (6.8-162.8)	27-64 (102.2-242.2)	44.5-107 (168.5-405.0)
Process Water, 10³ Kgal (10⁶L)	3-36 (11.4-136.3)	6-72 (22.7-272.5)	11-144 (41.6-545.1)	17-216 (64.4-817.6)	28-360 (105-1362.7)
Fuel Oil, Kgal (m³)	1-11 (0.9-10.0)	2-22 (1.8-20.0)	3-43 (2.7-39.0)	5-65 (4.5-59.0)	8-108 (7.3-97.9)
Coke, ton (MT)	2-20 (1.8-18.1)	3-40 (2.7-36.3)	6-80 (5.4-72.6)	9-120 (8.2-108.9)	16-200 (14.5-181.4)
Magnesium Oxide, ton (MT)	0.07-0.9 (0.06-.82)	0.1-1.8 (0.1-1.6)	0.3-3.5 (0.3-3.2)	0.4-5.3 (0.4-4.8)	0.7-8.8 (0.6-8.0)
H₂SO₄ (by-product), Kton (kt)	0.1-0.3 (0.38-1.14)	0.2-0.6 (0.76-2.3)	0.4-1.3 (1.5-4.9)	0.6-1.9 (2.3-7.2)	1.0-3.2 (3.79-12.1)
Liquid Waste, 10³ Kgal (10⁶L)	0.18-.7	0.5-1.4	0.8-2.7	1.3-4.0	2.1-6.6
Solid Waste, Kton (kt)	17-31 (18.7-29.4)	10-18 (11.0-19.8)	7-12 (7.7-13.2)	5-10 (5.5-11.0)	4-8 (4.4-8.8)
Operating Cost, $/ton (MT) Coal	108-127 (102.4-120.4)	65-77 (61.6-73.0)	42-50 (39.8-47.4)	34-40 (32.2-37.9)	27-31 (35.1-29.4)
Cents/10⁶ Btu (GJ) Input					

*Estimated by Battelle computerized model, based on full load boiler operation at 7008 hr/yr.

**Assuming 5 percent of installation cost (see also Table 10).

†Assuming 36 man-hr/day for magnesium oxide and citrate processes.

Table 23

Annual Cyclically Recurring Cost Elements of Citrate FGD Systems*

Boiler Size 10⁶ Btu/hr (GJ/hr)	25 (26.4)	50 (52.8)	100 (105.5)	150 (158.3)	250 (263.8)
Maintenance and Repair (10³ $)**	25-60	37-84	56-118	71-145	97-189
Labor, man-hr†	13140	13140	13140	13140	13140
Electricity (10⁶ kWh)	0.3-0.4	0.7-0.8	1.3-1.7	2.0-2.6	2.3-4.2
Cooling Water, 10³ Kgal (10⁶L)	24-82 (90-310)	49-164 (185-621)	98-329 (371-1245)	146-493 (553-1866)	244-822 (924-3112)
Process Water, 10³ Kgal (10⁶L)	2.7-4.6 (10.2-17.4)	5.3-9.3 (20.1-35.2)	11-18.6 (41.6-70.4)	16-28 (60.6-106)	26.6-46.5 (100.7-176.0)
Steam, Klb (MT)	120-1550 (54-703)	240-3100 (108-1406)	490-6200 (222-2812)	730-9300 (331-4218)	1200-15,500 (544-7031)
Citric Acid, ton (MT)	0.1-1.3 (0.1-1.2)	0.2-2.6 (0.18-2.4)	0.4-5.1 (0.36-4.6)	0.6-7.7 (0.5-7.0)	1.0-12.7 (0.9-11.5)
Hydrogen Sulfide, Kton (kt)	0.02-0.25 (0.02-0.23)	0.04-0.5 (0.04-0.5)	0.08-1.0 (0.07-0.9)	0.12-1.5 (0.11-1.4)	0.2-2.5 (0.2-2.3)
Soda Ash, Kton (kt)	0.002-0.02 (0.002-0.02)	0.003-0.04 (0.003-.04)	0.007-0.08 (0.007-0.08)	0.01-0.13 (0.01-0.12)	0.02-0.2 (0.02-0.2)
Sulfur (by-product), Kton (kt)	0.02-0.2 (0.02-0.2)	0.03-0.4 (0.03-0.4)	0.06-0.81 (0.06-0.73)	0.1-1.2 (0.1-1.1)	0.16-2.0 (0.15-1.8)
Liquid Waste, 10³ Kgal (10⁶L)	0.24-0.34 (0.91-1.29)	0.5-0.7 (1.9-2.6)	1.0-1.5 (3.8-5.7)	1.4-2.1 (5.3-7.9)	2.5-3.6 (9.5-13.6)
Solid Waste, Kton (kt)	0.2-0.8 (0.19-0.73)	0.5-1.5 (0.45-1.4)	0.9-3.0 (0.82-2.7)	1.4-4.4 (1.3-4.0)	2.3-7.3 (2.1-6.6)
Operating Cost, $/ton (MT) Coal	17-34 (18.7-37.5)	10-23 (11.0-25.4)	7-17 (7.7-18.7)	6-15 (6.6-16.5)	5-13 (5.5-14.3)
Cents/10⁶ Btu (GJ) Input	105-143 (99.5-135.5)	64-95 (60.7-90.0)	42-69 (39.8-65.4)	35-60 (33.2-56.9)	29-52 (27.5-49.3)

* Estimated by Battelle computerized model, based on full load boiler operation at 7008 hr/yr.
** Assuming 5 percent of installation cost (see also Table 10).
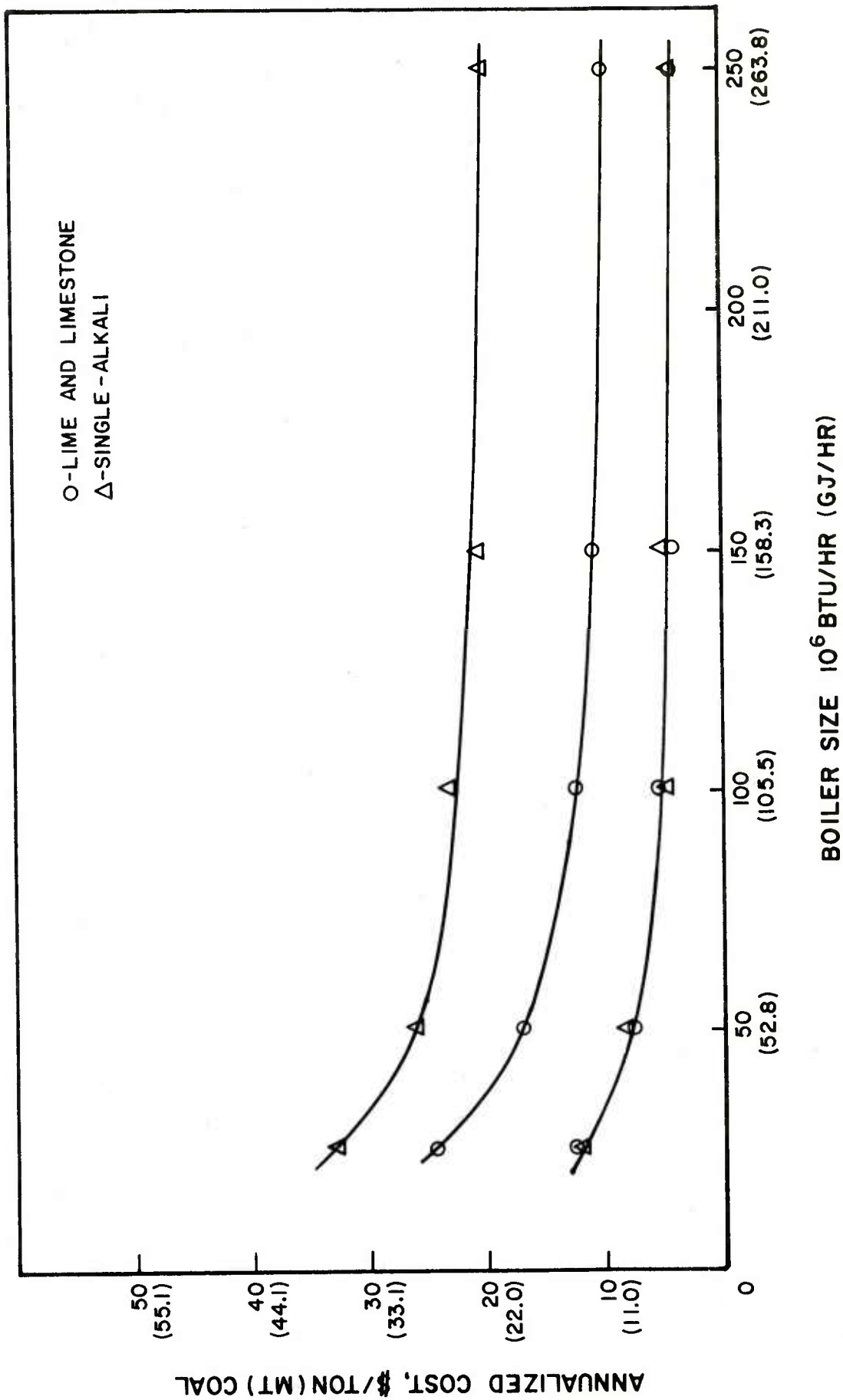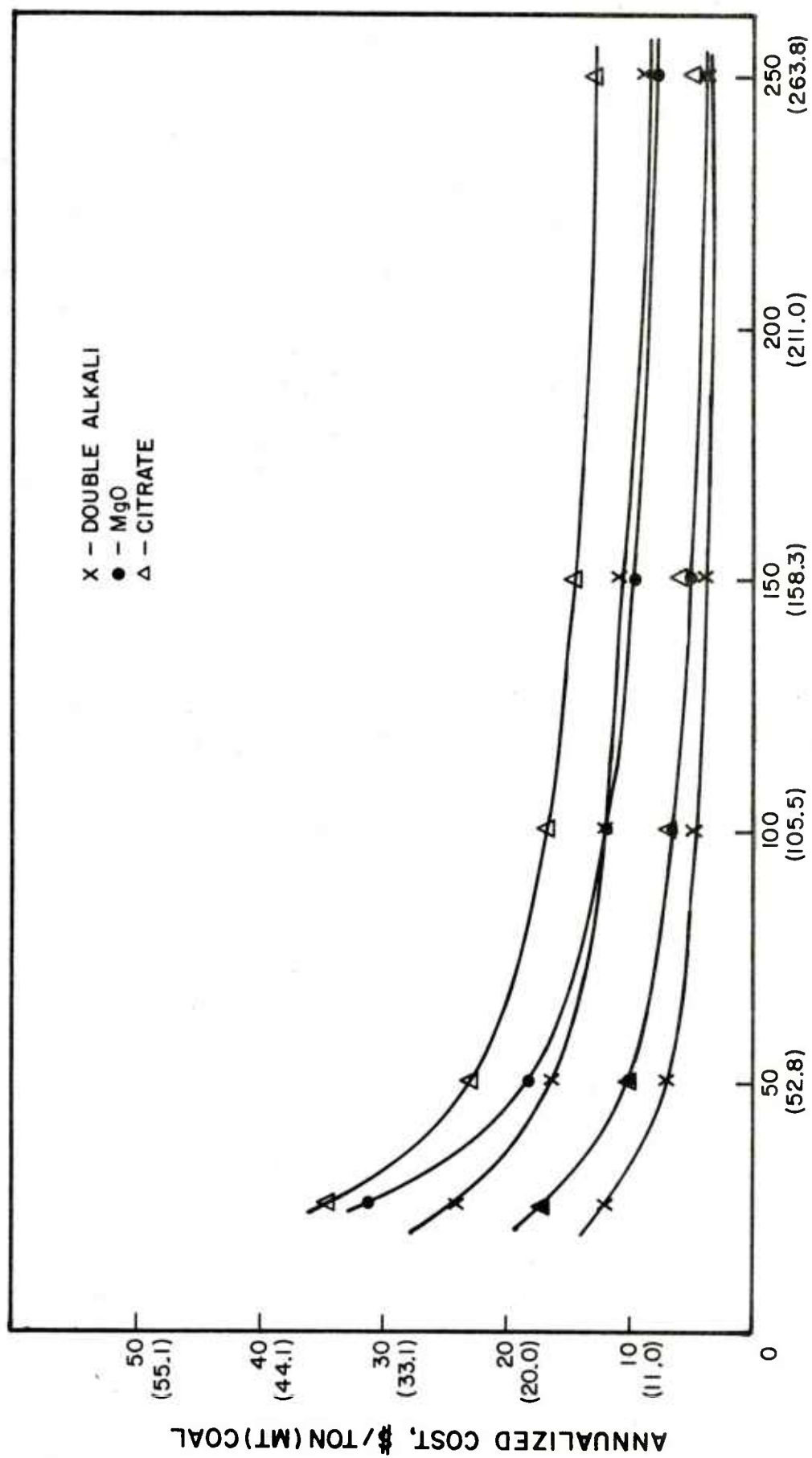† Assuming 36 man-hr/day for magnesium oxide and citrate processes.

Figure 23. Estimated operating cost of new FGD systems: lime, limestone, single-alkali (estimated ranges are between higher and lower bounds).

Figure 24. Estimated operating cost of new FGD systems: double-alkali, MgO, citrate (ranges shown as high and low limits).

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Wet Disposal or Ponding

For the wet disposal method, CERL estimated the capital cost (adjusted to October 1978) for establishing an environmentally acceptable pond; the estimate included all functions from excavation, embankment, lining, and fencing, to monitoring wells, piping, and instrumentation. The cost was scaled linearly from a data file prepared previously,[6] except a minimum cost of about $24,000 (equivalent to the costs for 1.6 km of three buried 1-in. (2.54 cm) pipes) was chosen for the pipeline requirement. The pipeline cost was scaled with an exponential factor of 0.4. This scaling procedure on pipeline cost, plus pond expenses, made the capital cost of wet disposal dependent on coal sulfur content, boiler size, and FGD process used. A 25 percent contingency was also applied to this capital cost estimation. Since the amount of solid-liquid mixed waste varies greatly with both the process employed and the sulfur content of the fuel, the investment costs of wet disposal also vary widely and are significant cost factors. The operating cost of waste ponding was estimated at about $1.13/ton ($1.25/MT) solid-liquid waste in all cases.

Dry Disposal or Landfill

For landfill disposal, capital costs (adjusted to October 1978) were scaled linearly from the computerized data file for all FGD processes except single-alkali. Since no pipeline is used for landfill, no exponential scaling was pertinent. This method of estimation, including a 25 percent contingency, resulted in a constant cost of about $6/ton ($6.61/MT) waste for all cases studied. Similarly, the operating cost for landfill practice, including transportation, was estimated at about $2.70/ton ($2.98/MT) waste for all processes.

[6] Battelle's Columbus Laboratories, Modeling of Economic and Energy Impacts of Flue Gas Desulfurization Systems, In-house Study (September 1978).

5 CONCLUSIONS AND RECOMMENDATIONS

This report has presented guidelines for selecting FGD systems (Chapter 4), and has analyzed and evaluated six existing FGD processes which could be used in Army coal-fired boilers:

1. The single-alkali process is advantageous because of its ease of operation, highly reactive agent, and freedom from plugging and sealing. However, waste disposal may cause problems.

2. Wet limestone scrubbing produces no by-product, can tolerate significant fly ash, and, of all systems, uses the least expensive reagent. However, the system's problems include the amount of waste produced, equipment corrosion, and low SO_2 removal efficiency (unless large amounts of limestone and very high liquid-to-gas ratios are used).

3. Lime scrubbers are simple to operate and can tolerate significant fly ash in the system. But the system produces large quantities of solid waste that must be disposed of, and equipment corrosion can be a problem.

4. Double-alkali systems present few scrubber plugging and scaling problems and allow a low liquid-to-gas ratio in the scrubber. However, the systems are complex and costly. At several industrial sites, there have been high particulate emissions from the scrubber due to misting.

5. The magnesia scrubbing process produces no major waste streams other than fly ash, is free from scaling and plugging problems, and offers high SO_2 removal efficiency. However, the system cannot tolerate fly ash, requires a high liquid-to-gas ratio in the scrubber, involves complex solids handling operations, and requires auxiliary fuel for the drying and calcining operations.

6. The citrate process presents no scaling and plugging problems and no waste disposal difficulties. However, the system requires many raw materials and a prescrubber to remove all residual fly ash.

Based on these conclusions, it is recommended that military installations limit their FGD considerations to the single-alkali, lime, and limestone processes at the present time. There is enough background and operational data on these processes to allow planning and engineering personnel to make reasonable assumptions about the systems' performance and appropriateness for use on Army boilers.

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